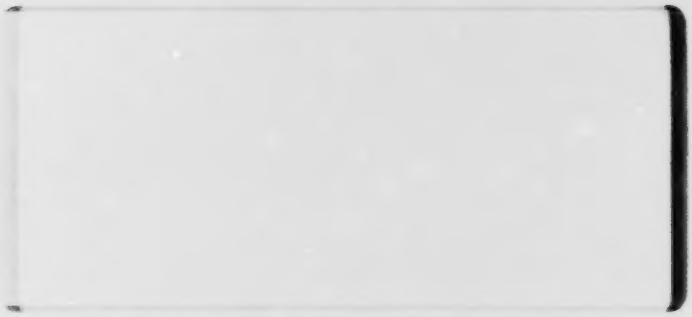

GULF CANADA LIMITED

RESEARCH AND DEVELOPMENT DEPARTMENT
SHERIDAN PARK, ONTARIO, CANADA

Final Report

DEVELOPMENT OF IMPROVED SYNTHETIC CRUDE PROCESSES AND THEIR IMPACT ON DISTILLATE AND OTHER PRODUCT QUALITIES

by

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TABLE OF CONTENTS

		Page No.
1.	SUMMARY	1
2.	INTRODUCTION	4
3.	EXPERIMENTAL	7
	3.1 Executed Work Statement	7
	3.2 P-12 and P-16 Microreactor Unit Description	11
	3.3 Sulphiding Procedure	21
	3.4 Data Evaluation and Product Characterization	23
4.	RESULTS AND DISCUSSION	24
	4.1 Crude Evaluations	25
	4.2 Hydrotreating Synthetic Crude Middle Distillate	27
	4.3 Hydrocracking Synthetic Crude Gas Oil	40
	4.4 Naphtha Disposition	49
5.	CONCLUSIONS	54
6.	RECOMMENDATIONS	56
	REFERENCES	58
	TABLES AND FIGURES	
	APPENDICES	

LIST OF FIGURES

Figure 2.1.1	Process Configuration
Figure 3.1.1	Project Overview
Figure 3.2.1	P-12 Hydrotreater Diagram
Figure 3.2.2	P-16 Hydrotreater Diagram
Figure 4.1.1	Crude Assay True Boiling Point Curve
Figure 4.2.1	Middle Distillate Aromatic Conversion vs. Reaction WABT
Figure 4.2.2	Middle Distillate TLP Molecular Weight vs. Reaction WABT
Figure 4.2.3	Hydrotreating Catalyst Ni Content vs. Middle Distillate Aromatic Conversion
Figure 4.2.4	Hydrotreating Catalyst Surface Area vs. Middle Distillate Aromatic Conversion
Figure 4.2.5	Middle Distillate Aromatic Conversion vs. Reaction WABT for Condition Optimization Runs
Figure 4.2.6	Middle Distillate Aromatic Conversion vs. Reaction Pressure
Figure 4.3.1	Hydrocracking Gas Oil Conversion vs. Reaction WABT
Figure 4.3.2	Middle Distillate Yield vs. Reaction Temperature
Figure 4.3.3	Middle Distillate Yield vs. Reaction Temperature for Optimization Runs

LIST OF TABLES

Table Number	Title of Table
3.4.1	Summary of CGSB Jet A Specifications
3.4.2	Summary of CGSB Jet B Specifications
3.4.3	Summary of CGSB Diesel Specifications
4.1.1	Crude Assay Data Summary
4.1.2	Crude Assay; Naphtha Reforming Yield Predictions
4.1.3	Crude Assay; Middle Distillate Mass Spectroscopy Analysis
4.1.4	Crude Assay; Gas Oil Mass Spectroscopy Analysis
4.1.5	Crude Assay; Syncrude, G.C. Distillation
4.2.1	Commercial Catalyst Data Summary
4.2.2-4.2.9	Middle Distillate Hydrotreating Catalyst Screening; Run Data Summaries
4.2.2	American Cyanamid HDN 30
4.2.3	Harshaw Ni4301E
4.2.4	Katalco NM506
4.2.5	Katalco NT550
4.2.6	Ketjen Armak KF840Q
4.2.7	Shell S324
4.2.8	Shell S454
4.2.9	Gulf Catalyst A
4.2.10-4.2.17	Middle Distillate Hydrotreating Catalyst Screening; Mass Spectroscopy Comparison
4.2.10	American Cyanamid HDN 30
4.2.11	Harshaw Ni4301E
4.2.12	Katalco NM506
4.2.13	Katalco NT550
4.2.14	Ketjen Armak KF840Q
4.2.15	Shell S324
4.2.16	Shell S454
4.2.17	Gulf Catalyst A

LIST OF TABLES Continued

Table Number	Title of Table
4.2.18-4.2.20	Middle Distillate Hydrotreating; Condition Optimization Run Data
4.2.18	Katalco NM506 10.3 MPa Runs
4.2.19	Katalco NM506 13.8 MPa Runs
4.2.20	Katalco NM506 17.2 MPa Runs
4.2.21	Middle Distillate Hydrotreating; Katalco NM 506 Mass Spectroscopy Data Summary
4.2.22	Middle Distillate Hydrotreating Production Run Data Summary
4.2.23	Middle Distillate Hydrotreating; Produced Naphtha Composition
4.3.1	Gas Oil Hydrocracking Catalyst Screening Runs; Katalco NT550
4.3.2	Gas Oil Hydrocracking Catalyst Screening Runs; Gulf Catalyst A
4.3.3-4.3.6	Gas Oil Hydrocracking and Condition Optimization Run Data Summaries
4.3.3	Gulf Catalyst A 13.8 MPa Runs
4.3.4	Gulf Catalyst A 17.2 MPa Runs
4.3.5	Gulf Catalyst A 20.7 MPa Runs
4.3.6	Gulf Catalyst A Supplementary Runs
4.3.7	Gas Oil Hydrocracking Production Run Data Summary Gulf Catalyst A
4.3.8	Gas Oil Hydrocracking Produced Naphtha Composition
4.3.9	Product Evaluation Data Summary - Jet B
4.3.10	Product Evaluation Data Summary - Jet A
4.3.11	Product Evaluation Data Summary - Diesel

NOTATION

API	American Petroleum Institute
ASTM	American Society for Testing and Materials
BP/BPCV	Back Pressure/Back Pressure Control Valve
CGSB	Canadian General Standards Board
DP/DPCV	Differential Pressure/Differential Pressure Control Valve
E	Extrudate
EOR	End of Run
EP	End Point of boiling temperature range
F1	Research octane number
F2	Motor octane number
FCCU	Fluidized Catalytic Cracking Unit
FIC	Flow Indicator Controller
FRC(V)	Flow Rate Controller (Valve)
GCD	Gas Chromatography Distillation
HC	Hydrocracked or Hydrocracker
HDN	Hydrodenitrogenation
HDS	Hydrodesulphurization
HT	Hydrotreated or Hydrotreater
IBP	Initial Boiling Point
L/L	Litres per litre (volumetric gas per liquid feed rate)
LIC	Level Indicator Controller
LHSV	Liquid Hourly Space Velocity
LRMS	Low Resolution Mass Spectroscopy
MPag	Mega Pascal (pressure) gauge
MON	Motor octane number
MW	Molecular Weight
P&ID	Process and Instrumentation Diagram
R+M	Research + Motor octane number
RON	Research octane number
RVP	Reid Vapour Pressure
SCR	Silicon Controlled Rectifier
SOR	Start of Run
SR	Straight Run
TBP	True Boiling Point
TLP	Total Liquid Product
TI(C)	Temperature Indicator (Controller)
TR	Temperature Recorder
VP	Vapour Pressure
WABT	Weight Average (Catalyst) Bed Temperature
WSIM	Water Separation Index

1. SUMMARY

In this work, four barrels of synthetic crude oil were obtained from Syncrude Canada Ltd. and processed on a pilot plant scale to determine product yields and qualities that could be obtained by using up to date refining processes and catalysts.

The catalyst screening involved hydrotreating syncrude middle distillate (149°C-385°C) with 8 commercial catalysts at repeated conditions to ascertain the relative performance of those selected. Two of these catalysts were also used in syncrude gas oil hydrocracking service. The objectives in hydrotreating catalyst performance were to achieve maximum aromatic saturation and minimum distillate cracking. A catalyst was chosen based on these performance criteria and was optimized for a middle distillate production run. The production run product was fractionated into jet and diesel fuels for evaluation. Similarly one of the two hydrocracking catalysts was chosen based on cracking activity and maximum middle distillate yield from the syncrude gas oil feedstock. The selected catalyst was also optimized and a production run was made to produce jet and diesel fuel for evaluation.

The production of fuel by hydrotreating and hydrocracking syncrude distillate components resulted in the following yields (distributed into four boiling ranges, IBP-15°C , 15-149°C, 149-385°C, 385°C+ on the the basis of one cubic metre of syncrude);

Distillation

1 m^3 whole syncrude,	IBP-15°C	23.3 litres
	15-149°C	157.0 litres
	149-385°C	604.2 litres
	385°C+	215.5 litres
	TOTAL	1000.0 litres

Hydrotreating

604.2 litres 149-385°C	IBP-15°C	2.6 litres
+ 95.9 m^3 H_2	15-149°C	47.1 litres
	149-385°C	597.8 litres
	TOTAL	647.5 litres

Hydrocracking

215.5 litres 385°C+	IBP-15°C	20.1 litres
+ 48.5 m^3 H_2	15-149°C	31.6 litres
	149-385°C	145.4 litres
	385°C+	58.7 litres
	TOTAL	255.8 litres

Combined Processing

23.3 litres	IBP-15°C ---	IBP-15°C	46.0 litres
157.0 litres	15-149°C ---	15-149°C	235.7 litres
604.2 litres	149-385°C --	149-385°C	743.2 litres
215.5 litres	385°C+ ---	385°C+	58.7 litres
144.4 m^3 H_2	---		
		TOTAL	1083.5 litres

Three sets of fuels were evaluated. Two sets were comprised of products of the hydrotreating and hydrocracking operations as stand alone processes. This enabled the assessment of the two

processes independently. The other set of fuels was comprised of products of the hydrotreating and hydrocracking operations blended together according to the balance found in the original, whole syncrude to form the basic fuel pools; Jet A, Jet B and Diesel.

Overall, a high yield of middle distillate (74.3 vol% on crude) was demonstrated. The diesel fuels made were of good cetane value (45-46), had good low pour point qualities and met all criteria for diesel fuel specifications. The jet fuels produced met or exceeded the respective quality criteria.

The severity of operation utilized in these processing schemes have mixed cost/benefit results. The hydrotreating operation, modelled by this work, could be reduced in severity (e.g. increased space velocity) to meet minimum CGSB specifications and achieve reduced costs of operation. Hydrotreating syncrude middle distillate can also be traded off against ignition and combustion property improvement by the use of additives and/or against blending with conventional middle distillate. The deep cut middle distillate used in this work would not likely be used in these last scenarios due to poor flow properties of the high end point material and the middle distillate boiling range would have to be narrowed resulting in a yield reduction.

The hydrocracking operation is a unique process configuration, and in order to achieve significant results, severity similar to that used in this work is necessary. The effects of high severity on the catalyst, such as deactivation, were not measured. Likewise catalyst poisoning was not evaluated. The production of middle distillate from hydrocracked gas oil is an important route to increased middle distillate yield from synthetic crude.

2. INTRODUCTION

The Canadian petroleum industry has been experiencing some uncertainty in direction. This has been predicated by recovery from the oil shock of the seventies, shifting crude slates and end user product consumption. A challenge to be faced in this environment is the integration of synthetic crudes into refinery processing to produce quality fuel and oil products economically.

The North American oil industry has seen and continues to see a number of new and different types of crude stocks which include significant new supplies of heavy and synthetic crudes. At the same time product demand is shifting emphasis to middle distillate fuels as consumers make efficacy gains on gasoline consumption. Growth in certain areas of the economy, such as resource and transport industries, will maintain the demand for middle distillates thereby changing traditional gasoline/diesel ratios.

The oil refining business, in responding to economic conditions has increased hydrogenation capacity, increased hydrocracking capacity and is involved in vacuum tower bottoms upgrading. These were introduced to allow utilization of a wider range of crudes and provide for an overall increase in yields. This is a strategy to improve flexibility and hopefully refinery profit margins. Canadian refiners, in particular, have invested in yield improvement processes despite the general downsizing of the refining business.

The penetration of synthetic crude into the Canadian crude supply is significant and currently represents close to 10% of the national crude production. Synthetic crude, commercially produced at Syncrude Canada Ltd. and Suncor Inc. tar sands projects, presents the purchasing refiner with a mixed benefit/detriment feedstock. The low sulphur, low salt and bottomless characteristics of synthetic crude should make it of good value. However, these are offset by the high aromaticity of the middle distillate and gas oil fractions.

The conventional refinery operation typically blends synthetic crude into the total crude charge and as such foregoes the efficiency of dedicated processing. The recent installation and start up of synthetic crude specific units (Shell Scotford, Esso Strathcona, and Gulf Edmonton) mark improvements in synthetic

crude utilization. However, even these have been developed based on conventional petroleum relationships and extensions of existing technology. Therefore, there continues to exist a need to evaluate syncrude processing and product technology to enable the efficient utilization of the resource.

The improvements in technology necessary to better utilize synthetic crude and to produce conventional quality products includes:

- evaluation of syncrude naphtha as reformer feed and comparison with conventional naphtha feeds
- processing for improving ignition and combustion properties of diesel fuels and kerosenes derived from synthetic crude
- improvement of synthetic gas oils for use in catalytic crackers
- processing of synthetic gas oil for direct conversion to usable transportation fuel by hydrocracking.

It is evident from the above that a major element in process development for synfuels production is catalyst performance. Also, process configuration and condition optimization are keys to the improvement of products from synthetic crudes.

The work contained herein addresses the product and yield improvements achievable in available hydrotreating and hydrocracking processes applied to middle distillate and gas oil fractions derived from synthetic crude.* The objectives of this work were:

- Predict product qualities and yields achievable through improved refining processes for a synthetic crude and assess related product costs.
- Establish the performance of existing catalysts, with a view to identifying the needs for catalyst development.
- Obtain a clearer understanding of the basic petroleum chemistry of relevant hydrogenation and cracking processes, by means of a thorough characterization of the feeds and products.

The benefit of such work is to

- show the route to improved near and long term strategic supply of quality fuel products from a typical synthetic crude,
- demonstrate commercial catalysts in applications to synfuel processing and show areas of needed improvement,

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* Synthetic Crude produced by fluidcoking of Athabasca bitumen.

. contribute to the technology base necessary for progression in synthetic crude utilization and optimization of this resource.

The main objectives of this project were:

- selection of a typical, commercial synthetic crude
- feedstock preparation and characterization
- performance screening of a field of commercial catalysts in hydrotreating middle distillate and hydrocracking gas oil
- performance optimization of the best hydrotreating and best hydrocracking catalyst
- hydrotreating and hydrocracking production runs
- evaluation of the gains in product quality

The overall processing configuration that was simulated in this project is shown in Figure 2.1.1. The configuration represents a simplification of a typical commercial synthetic crude system. As such, the front end has only an atmospheric fractionator and no provision for a vacuum distillation owing to the bottomless composition of synthetic crude. The flow sheet shows the inclusion of the gas oil hydrocracker which represents a maximum middle distillate operation. The gas oil would alternately have to go to a FCCU if the hydrocracker was not needed for middle distillate production. In the case of Gulf's Edmonton synthetic crude unit the hydrocracker is actually a dual purpose hydrotreater/ hydrocracker and reduces the number of pieces of equipment accordingly.

The results of the technical program are the subject of the following report.

3. EXPERIMENTAL

3.1 Executed Work Statement

The work in this project was divided into nine principal tasks involving three basic areas of investigation and two modes of operation. The project, however, evolved somewhat as the work progressed and the piloting runs were changed to accommodate operating conditions of interest. An overview of the work performed is illustrated in Figure 3.1.1 which is a block diagram of the program components showing the flow of work. The nine tasks of the work statement (as performed) are described below.

Task No. 1 Distillations and Crude Assay

The distillation of three and a half barrels of typical Athabasca syncrude into three fractions was performed which included a naphtha fraction, IBP-149°C; a wide range middle distillate, 149-385°C; and a gas oil, 385°C+. This distribution of boiling range was to reflect the widening of middle distillate boiling points generally, projecting current trends. The volumes distilled were necessary for the completion of all the catalyst screening and optimizing work as well as to provide sufficient feed for production runs and to supply product testing quantities.

A crude assay was performed to characterize the feedstocks and establish the representivity of the crude sample. The feedstock analyses provided baseline data on which the performance of subsequent hydrotreatment could be assessed. This was done to improve the understanding of the chemistry involved in hydrotreatment of these synthetic stocks.

Task No. 2 Catalyst Screening Runs on the P-16 Microreactor

Preliminary screening of eight middle distillate hydrotreating catalysts and two gas oil hydrocracking catalysts was carried out. The catalysts were selected on the basis of known performance criteria and properties such as composition. The selection was done jointly by the Scientific Authority and Gulf.

The hydrotreating catalysts were screened as to their relative capability to remove trace sulphur and nitrogen and to determine

the degree of aromatic conversion achieved. The change in product molecular composition was determined by mass spectroscopy. Similarly the performance of the two hydrocracking catalysts was measured by observing the change in composition of the products and included mass spectrometric analysis for the following molecular types: Paraffins, Cycloparaffins (single, double and polycyclic), and Aromatics (alkyl benzenes, tetralins, dinaphthenebenzenes and naphthalenes).

Catalyst extrudates were tested whole in the P-16 pilot plant microreactor, which produced between two and four litres per day of liquid product. After sulphiding and preliminary de-edging of each of the catalysts a series of three runs was made. Each hydrotreating catalyst was screened at 13.8 MPag pressure and 1.0 LHSV for three different temperatures: 350, 385 and 420°C. This provided a common basis comparison and a profile of activity for each catalyst. The hydrocracking catalysts were screened at 20.7 MPag pressure, 0.5 LHSV and temperatures of 385, 405, and 420°C.

Task No. 3 Fixed Bed, Plug Flow Hydrotreating Reactor Pilot Runs

The runs performed in Task 3 were used to establish optimum conditions of temperature and pressure. The runs consisted of 3 pressures (10.3, 13.8, and 17.2 MPag) x 3 temperatures (335, 360 and 385°C, plus supplementals @ 410°C) to map out catalyst performance over a range of treatment conditions; a total of 11 pilot runs on the P-12 hydrotreater were carried out. A constant LHSV of 1.0 was used and all other conditions were maintained constant for comparison purposes.

The products of the runs were analyzed for specific gravities, light ends, boiling point distribution, sulphur and nitrogen, and compositional analysis by mass spectroscopy. A material balance on the hydrotreater system for each run was performed. The results were used to determine optimum operating condition in conjunction with advice from the Scientific Authority.

Task No. 4 Fixed Bed, Plug Flow Hydrocracking Reactor Pilot Runs

The selected hydrocracking catalyst was loaded in the P-16 pilot reactor as approved by the Scientific Authority for optimization

runs at 3 conditions of pressure (13.8, 17.2, and 20.7 MPag) each with a range of three temperatures (360, 385 and 400°C). At the most effective pressure, 2 further runs were performed at two more temperatures (410 and 420°C). The products were analyzed for specific gravity, light ends, boiling point distribution, sulphur and nitrogen and molecular composition. After each run, a mass balance was performed and these results combined with the analytical data were used to determine the optimum treatment condition for the subsequent production run.

Task No. 5 Calculation of Ultimate Product Yields Using Models for Naphtha and Pilot Plant Results for Distillate and Gas Oil

The yields of products from Tasks 3 and 4 were calculated by using material balances determined from the pilot optimization runs. The limited quantities of naphtha produced in the optimization runs precluded adequate analyses for reformer predictions. The naphthas produced in the final production runs were characterized and reformer predictions were performed for these runs. These calculated estimates of feeds and yields of products were used in the determination of optimum operating conditions.

Task No. 6 Pilot Plant Hydrotreating Production Run

The assessment of optimum operating conditions for hydrotreating syncrude middle distillate were jointly reviewed and conditions for the production run were set with the approval of the Scientific Authority. The P-12 pilot hydrotreater was used and the conditions of 17.2 MPag, 385°C, and 1 LHSV were chosen. The catalyst was aged in preparation for the run such that the activity was relatively constant for the entire production run.

The product was analyzed for comparison with the optimization runs and pilot unit material balances for the run were also completed. The total liquid product was subsequently distilled into the original boiling ranges and portions of the middle distillate and naphtha fractions were cut into blending components for fuel testing.

Task No. 7 Pilot Plant Hydrocracking Production Run

The P-16 pilot plant hydrotreater was used to hydrocrack the syncrude 385°C+ gas oil at conditions of 20.7 MPag, 405°C, and 0.5 LHSV. The catalyst was prepared and de-edged to stabilize activity and run to hydrocrack the prepared feedstock. The product was inspected for the same properties as the optimization run products and was distilled into the original boiling point ranges. The middle distillate and naphtha fractions were further distilled to produce the blending components for Task No. 8.

Task No. 8 Evaluation of Production Run Middle Distillates and Hydrocracked Gas Oil as Potential Transportation Fuels

The evaluation and testing of fractionated products from the two production runs was done on formulated fuels including:

a) Motor Gasoline
b) Aviation Turbine Fuels Jet A and Jet B
c) Diesel Fuel

The middle distillate fuel products were tested according to CGSB specifications including combustion properties, low temperature operability, thermal stability and oxidation stability. Motor gasoline was evaluated based on reformer predictions and raw naphtha property testing including motor octane.

Task No. 9 Compilation of Monthly and Final Reports and Presentation

This task was the preparation of monthly progress reports and the composition of the final report.

A presentation of mid-term results in the project was given in May (1984).

3.2 P-12 and P-16 Microreactor Unit Descriptions

3.2.1 Introduction

The units utilized in this study were both of a similar, medium-small scale. The P-12 unit had a catalyst bed volume of 320 cc whereas the P-16 unit was nominally sized at 170 cc. The principal difference between the two units, apart from the size, was that the P-12 had recycle gas capability and the P-16 unit was designed for once-through gas operation. The once-through reaction system permitted better run reproducibility, quicker line out and stabilization as well as ease of operation.

Hence, the P-16 reactor lent itself well to catalyst screening, did not require the extra line out time necessary to bring recycle gas to equilibrium, and was very suitable for Task #2. However, these advantages were also applicable to condition optimization and production run situations. The recycle capability of the P-12 unit was provided only to simulate commercial operations, but was not essential for the work set out in this project where the relative performance of catalysts was measured.

The following description of the hydrotreating process is particularly for the P-12 unit. The P-16 unit was largely similar in design and layout. Figures 3.2.1 and 3.2.2 show P&ID for each reactor system respectively. The P-16 reactor had less heaters and controllers because the intended application was restricted to distillate and gas oil treatment.

3.2.2 Description of the Hydrotreating Process

The 5-gallon charge tanks were filled from the feed drum using 14 kPag maximum nitrogen pressure. From the feed tanks which were kept under 5" H_2O gauge nitrogen pressure, feed passed through a filter to the suction manifold of the feed pumps P1 and P2. The pumps were equipped with 27.6 MPag safety discs, so that if a disc ruptured the charge would circulate from pump discharge to pump suction. A check valve was provided to prevent back flow from the reactor should a disc rupture.

The feed joined the recycle hydrogen gas stream and the mixture passed through an external preheater, an internal spiral screw preheater situated in the top of the reactor and then down through the catalyst bed where reaction occurred.

The gaseous products which left the reactor passed through the water-cooled condenser and into the high pressure separator where the gas and liquid phases were separated. Liquid in excess of that required to maintain the specified level in the separator passed through a filter, a control valve LCV1 and a Jerguson gauge glass fitted with a drip tube to observe the flow.

The gas from the high pressure separator passed through the recycle gas driers (which could be by-passed if desired) then through a surge tank and filter and into the suction manifold of the recycle compressor. Excess pumping capacity from the discharge manifold passed through a research control valve DPCV1 and returned to the suction manifold. The remainder of the gas from the discharge manifold passed through a surge tank and split into two streams, the recycle gas and the reactor gas. Reactor gas in excess of that necessary to hold the desired hydrogen content or reactor pressure constant was bled off through a BPCV2 control valve and then to valves which forced part through a gas chromatograph sample manifold. The gas from the chromatograph rejoined with the balance of the stream and passed through a manual sampling loop and the wet test meter before going to vent.

The recycle gas passed through a filter, research control valve FRCV1, then through a McAfee displacement gauge manifold where the gas rate was measured. It then passed through a metering valve and a check valve before joining the feed at the inlet to the preheater.

3.2.3 Equipment

External Reactor Preheater

The preheater was formed by winding 1.4 m of 64 mm O.D. x 0.9 mm wall tubing around a 0.3 m length of 620 mm Sch. 40 pipe closed at one end. The pipe was also wound with a 2000 watt calrod element and the tube and calrod were thermonized together and to the pipe. A skin thermowell was welded on to the pipe near the lower end. The grade of Thermon used here was good to only 650°C. The transfer line and the reactor head were wound with a heating tape. The space between the preheater and the furnace top was enclosed in a metal box and filled with loose vermiculite.

Reactor P-12

Internal Dimensions were 1.2 m x 0.22 mm diameter. An internal preheater screwed into the top reactor head. The main one was 0.3 m long. This could be extended in length by the use of either of two preheater extensions 0.26 m or 0.14 m in length. Reactor Deadmen also were available in 3 lengths, 0.48 m, 0.40 and 0.25 m to take up space in the bottom of the reactor. Variation in the deadmen and preheater sizes permitted variations in size and location of catalyst beds. Suitable combinations for the various size beds were as follows:

Bed Size	Preheater Size	Deadman Size
0.13 m	0.46 m	0.48 m
0.20 m	0.46 m	0.40 m
0.26 m	0.46 m	0.40 m
0.41 m	0.46 m	0.25 m
0.71 m	0.13 m	0.25 m
0.86 m	0.13 m	None

The location of the top of the 0.15 m heating zone was 0.30 m above the reactor bottom. The top of the 0.46 m heating zone was therefore 0.76 m above the reactor bottom.

Reactor P-16

The reactor itself was 1.5 m long by 14.2 mm I.D. A preheater was provided upstream of the reaction system and all reactor heaters were externally mounted Chronolox units. The nominal bed size was 170 cc which corresponded to a 1.2 m bed length. The general operation was the same as the P-12 reactor with the exception of the recycle system which does not exist for P-16.

Reactor Furnace

Basically this was a bronze block 120 mm O.D., with a hole 38 mm O.D. in the centre, in which the 32 mm O.D. reactor was inserted. The block was divided into 3 sections insulated from each other by asbestos paper. The 3 zones were wound with heating elements as follows:

Section	Top	Middle	Bottom
Length	0.3 m	0.46 m	0.5 m
Wattage	2000	1000	1000

A single hole was drilled through the 3 sections to carry thermocouples and the whole was mounted in insulation in a steel case. The bronze block provided a heat sink to permit easier temperature control.

Reactor Products Condenser

This was made by winding a 3 m length of 6.4 mm O.D. x 1.6 mm wall tubing along with a similar length of copper tubing into a tight coil. The two coils were clamped together and covered with Thermon. A similar method of construction was used for the stabilizer bottoms cooler - however in this case only 1.22 m of tubing is required.

High Pressure Separator

This vessel had a 90 mm diam. x 250 mm (inside dimension) upper section, with a 0.32 mm long section of 13 mm Sch. 80 pipe welded on the bottom. This lower "dip-leg" had a volume of approximately 60 mL and was intended to minimize the effect of variations in separator level on unit hold up. The upper section was fitted with a 50 mm deep section of Metex extra dense demister pad. The vessel was wound with a copper tubing cooking coil which was thermonized to the wall and then covered with foil insulation. Cooling water on the reactor outlet cooler plus the separator jacket gave temperatures of 7-10°C in the separator. Higher temperatures (35-45°C) were achieved by the use of a heating tape controlled by a powerstat, with or without cooling water.

Compressor Surge Tanks

One of these vessels was provided on both suction and discharge of the compressors, in order to dampen pressure surges resulting from the diaphragm compressors and to provide an increased gas volume to smooth gas flow fluctuation.

The vessels used were the Aminco 65 mm O.D. micro series reaction vessels. These vessels were 38 mm diam. and 0.27 m long (inside measurement). The vessels were supported vertically with the closure downward. The bottom connection was fitted with a drain.

Gas entry was through the side connection at the bottom and the exit at the centre top. A 50 mm thick Metex extra dense demister pad was placed about 2/3 the distance from the bottom.

McAfee Gauges

These were positive displacement measuring devices for measurement of flow under high pressure on both the make-up and recycle gas streams.

3.2.4 Instrumentation and Control

3.2.4.A Temperatures and Heaters

1) External Reactor Preheater and Transfer Line

The reactor preheater temperature control TIC1 was set to control the gas inlet temperature to the reactor up to a maximum skin temperature of 540°C, the preheat being obtained from a 2000 watt 110 volt calrod. The power to the calrod was controlled by a SCR, as was the power to the windings on the transfer line between the preheater and the reactor. The preheater skin temperature was recorded while the transfer line temperature was indicated. The P-16 unit operated with similar equipment and control features.

2) Reactor Furnace Block

The P-12 unit block was in three sections, each being wound with adjustable 110 volt heaters. Each heater was provided with a thermocouple point located in the furnace block at the middle of the windings. There were two thermocouples at each point, one for TIC and the other for TR. The top section 2000 watt, was 0.3 m long controlled by TIC2. The mid section 1000 watt was 0.46 m long, controlled by TIC3. The lower section 1000 watt was 0.15 m long and controlled by TIC4. Maximum power to each heater was controlled by individual SCR's. Maximum block temperature was limited to 563°C. The P-16 unit had two heaters in the furnace block each with 1500 watts capacity.

3) Catalyst Bed

A thermowell extended from the reactor base through the full length of the catalyst bed and housed an eight-point thermocouple bundle T12-9. In addition, four of the points were recorded by TR5-8. If any temperature on the recorder was above a specified point, an alarm bell rang. A number of thermocouple bundles were available with different spacings for various sizes of catalyst bed. The P-12 catalyst bed temperature was taken as the average of the 18 TI's. The P-16 catalyst bed temperature was taken as the average of 4 TI's.

4) Condenser

The product line was coiled and traced, the two coils thermoned with standard thermon. The trace could carry steam or water.

5) High Pressure Separator

The high pressure separator was cold water traced and thermoned with standard Thermon. The skin temperature of the liquid level was measured on the multipoint indicator at TI-21. In the case of P-16 unit, the high pressure separator was a special Jerguson gauge designed for 20.7 MPag operation.

3.2.4.B Pressures

1) Reactor

The reactor pressure was controlled by Annin control valve BPCV2, actuated by a Fischer and Porter indicating controller. The gas tap was taken at the suction surge tank to the recycle gas compressors. When the hydrogenation was taking place, make-up gas was admitted on pressure control, whereas on reforming gas was discharged on pressure control. The P-16 pressure control was achieved by a Tescom back pressure regulator.

2) Compressor Discharge

Excess compressor pumping capacity was controlled by DPCV-1 which opened to allow gas to circulate from compressor discharge to compressor suction manifolds. DPCV-1 was actuated by

a Moore pneumatic controller which received a signal from a D.P. cell tapped across recycle gas rate control valve FRCV-1. A differential relief valve, 0.7 MPag, was also fitted between compressor discharge and suction manifold. The P-16 unit received hydrogen from the pilot plant high pressure supply system.

3.2.4.C Flow Rates

1) Feed Charge Rate

The feed charge rate was set by controlling the stroke length of either of the two charge pumps. Ideally, a minimum of 50% full stroke length was used for constant flow rates as these pumps did not meter accurately at lower stroke settings.

2) Recycle Gas Rate (P-12 only)

The recycle gas rate was under cascade control whereby a needle valve was used to create a pressure drop across a DP cell, and FRCV-1 and DPCV-1 then regulated the rate of flow to keep this pressure drop constant. The needle was set so that FRC-1 operated in the top 50% of the range.

FRCV-1 which was on the recycle gas line was operated by a Fischer Porter controller which received a signal from the DP cell across the needle valve. A low flow alarm switch was situated on the output of this DP cell.

Any movement of FRCV-1 was detected by a DP cell tapped across it and a signal was transmitted to a Moore 50 pneumatic controller which operated DPCV-1 on the compressor by-pass line.

Any fluctuations in flow were detected first by the DP cell across FRCV-1 and therefore DPCV-1 was used for rough control to prevent sudden surges and acted before FRCV-1.

The gas rate was measured by a positive displacement McAfee gauge located downstream of FRCV-1.

A constant rate was maintained if FRCV-1 and DPCV-1 were on control. Adjustments of a small degree were usually made by

adjustment of the controller set point. If larger adjustments were required the compressor stroke was altered and alterations made in the variable orifice needle valve 78 setting, to insure that all instruments were on control and a smooth record produced on FRC-1 while maintaining a uniform McAfee Cycle time. Adjustments to proportional band and reset could be required.

3) Make-up Gas Rate

This was measured by the make-up McAfee gauge, a duplicate of the recycle McAfee but with a closer reed switch spacing to allow for the lower flow.

The P-16 unit hydrogen input was measured by an orifice meter, calibrated before each run.

3.2.4.D Levels

1) Separator

The liquid level in the separator was maintained by the research control valve LCV-1 on the liquid line from the separator to the stabilizer. Taps from the gas and liquid sides of the separator were connected to a DP cell. The impulse was transmitted by this cell to the Moore controller, which regulated LCV-1 . The level was maintained just at the top of the separator dip leg. A low level alarm switch was located on the output of the DP cell.

3.2.4.F Analyzer

1) Gas Chromatograph

This was provided with a special manifold so that it could alternately sample the reactor gas and stabilizer overhead gas.

3.2.5 Experimental Procedure

Pressure Testing

The unit was thoroughly pressure tested before starting up on any run after a shut-down or when the unit had been opened for any reason by breaking high pressure fittings. The pressure test included the liquid system from the discharge of the feed pumps P-1 and P-2, to the separator liquid level control valve LCV-1, and included all of the recycle gas system and gas product line to reactor pressure control valve BPCV-2. The make-up gas system, including the make-up McAfee was also pressure tested. The P-16 unit was pressure tested in its entirety.

1. Cold Nitrogen Test

The unit was purged thoroughly with nitrogen making sure that the gas flow was downward through catalyst bed. The unit was pressured to 1.4 MPag and depressurized. This was repeated twice. The McAfee gauges were open to the system and also nitrogen purged.

The unit was pressurized to 10.3 MPag with nitrogen. This was done with compressor discharge pressure control valve DPCV-1 open.

2. Cold Hydrogen Test

When a satisfactory nitrogen test had been obtained, the unit was depressurized through the vent line and purged with hydrogen. It was pressurized with hydrogen to 1.4 MPag through the make-up system and then the pressure was released to vent. This procedure was repeated twice. Under at least 1.8 MPag the hydrogen was circulated using both compressors. Hydrogen was continuously admitted and discarded to vent with BPCV-2 on control. McAfee gauges were operated. Then the compressors were shut down and gas inlets and outlets were blocked. During the above procedure the BPC-2 controller action was set in position to release gas on pressure rise if reforming operation was to be done, or to admit gas on pressure drop if hydrogenation was to be done. In the latter case, gas was vented manually.

The unit was pressurized to 10.3 MPag with hydrogen, with compressors shut down, and as many leaks as possible were corrected.

This test was for low pressure work only i.e. <6.9 MPag. The Barton Gas Detector was used for locating hydrogen leaks.

3.2.6 Start-Up Procedure

a) After a cold hydrogen test had been obtained, BPCV-2 was again activated bringing pressure to run condition. The hydrogen circulation was started through the unit at the required flow rate and the furnace block temperatures raised at a rate of 28°C (max.) per hour.

   The catalyst bed temperature was monitored by a single "probe" thermocouple while the reactor held at a steady temperature of 200°C. The temperature was recorded by this thermocouple at each of the points which was indicated by a multicouple. The probe thermocouple was then replaced by the multicouple in the thermowell and the temperatures checked against those previously recorded. If the temperatures agreed to $\pm$1°C then the multicouple was correctly located and accurate temperatures were being indicated.

b) Water flow was started through all coolers.

c) Catalyst bed temperatures were reduced to slightly lower than the desired temperature for the subsequent run.

d) The feed pump was primed, discharging feed to a can and was then set for the desired initial rate using a calibrated gauge glass.

e) The feed was started into the unit, and the feed rate was checked by pumping from the calibrated gauge glass. Initial feed rates were about half the normal. After pressure and temperature steadied out, the feed rate and reactor temperature were increased to specified values.

f) Temperatures, pressures, levels and rates were re-checked as often as possible during the line-out period, recording hourly on a log sheet.

g) The gas recycle rate was readjusted to the desired flow using the actual recycle gas composition.

h) The product vessel was charged hourly and the API recorded.

i) Hourly gas analyses were taken to determine if specified conditions were met and as an aid in assessing when operation had become steady. When operation was steady the run was begun and product collected for specified intervals.

3.3 Catalyst Presulphiding

The sulphiding of the various catalysts was carried out according to a standard procedure, which is outlined as follows:

- The reactor was charged with sieved and dried catalyst in the arrangement specified for normal charge.
- The reactor was subsequently pressurized and leak tested using H_2 @ 27.6 MPag such that pressure loss was no greater than 14 kPa per hour.
- The reactor was warmed up to 149°C and H_2 was purged or circulated to dry the catalyst until no water was apparent in the line traps following a final leak check.
- Syncrude middle distillate was used as the presulphiding feed and was spiked with CS_2 to the equivalent of 6000 ppm sulphur. After the feed flow had been started, the warm-up cycle was commenced at 28°C/h until the reactor reached 300°C (the first sulphiding temperature). The gas rate was set at 890 L/L with a minimum of 93% H_2 composition.
- Presulphiding was carried out at 20.7 MPag pressure and 1.0 LHSV and these conditions were maintained until the prescribed amount of sulphur had been passed over the catalyst. The time required to complete the operation was estimated as given below. As an indication of when the sulphiding was complete, a breakthrough of H_2S in the recycle gas was monitored for. Each H_2S breakthrough was followed by a temperature increase of 30°C until 390°C was reached.
- The manufacturer's specifications were used to determine the percentage of sulphur added to the catalysts.
- The H_2S concentration in the effluent gas was monitored for 2 hours after the final breakthrough to ensure that sulfiding was complete.

Example Calculation (e.g. Katalco NM 506)

Feedrate = catalyst charge x LHSV
= 0.323 L x 1.0
= 0.323 L/h

Weight of Catalyst = catalyst charge x bulk density
= 0.323 x 920 g/L
= 297.2 g

NM 506 Metals = 5.3 wt% Ni x 297.2 g = 15.8 g = 0.27 moles Ni

= 18.0 wt% Mo x 297.2 g = 53.5 g = 0.56 moles Mo

$3Ni + 1S_2 \quad 1Ni_3S_2$

$1Mo + 1S_2 \quad 1MoS_2$

Moles S_2 required = $\frac{0.27}{3} + 0.56 = 0.65$

= 41.7 g S_2

CS_2 Required = $\frac{MW\ CS_2}{MW\ S_2}$ x sulphur required

= $\frac{76.131}{64.128}$ x 41.7

= 49.5 g (@ 6000 CS_2 spike; ppm in feed)

Quantity of Feed Required = CS_2 required x 1/spike concentration

= 49.5 x $\frac{1000}{6}$

= 8.3 kg Feed

Hours of Sulphiding = Feed required x 1/sg Feed x 1/LHSV

= 8.2 x $\frac{1}{0.8644}$ x $\frac{1}{0.323}$

= 29.7 hours

3.4 Details of the Data Evaluation and Product Characterization

The raw and processed data was reviewed thoroughly with the Scientific Authority through monthly reports, visits, and in an interim presentation at the EMR offices (May of 1984).

The data was generated by pilot plant operations consisting of fractionation, catalytic hydrogenation and hydrocracking. In addition, data was generated from analytical petroleum type tests on the synthetic crude feed, component products, and blended products.

Regarding evaluations of the products, and product characterization, these were defined in Schedule "A", page 4 of 5 in the contract Agreement. The latest National Standards of Canada specifications for evaluating the fuels properties were used. These are set by the Canadian General Standards Board. In general, properties of fuels are determined by laboratory tests set and standardized by a number of organizations. The main ones are:

ASTM - American Society for Testing and Materials
IP - Institute of Petroleum
ISO - International Organization for Standardization
CGSB - Canadian General Standards Board

The national standard specifications for Canadian diesel fuels and jet fuels are included in this report, in Tables 3.4.1 - 3.4.3.

4. RESULTS AND DISCUSSION

Task 1 of the work statement provided the characterized feedstocks for the treatment studies plus an assay of current Syncrude Canada Ltd. product. Three and a half drums of synthetic crude oil were taken from the Edmonton-Fort McMurray pipeline on February 17, 1983 while the Syncrude plant was running at typical quality production. The crude assay performed revealed that this sample was quite representative of other syncrudes produced over the life of the Syncrude project and was indeed typical as the following data shows.

The balance of the characterized feedstocks were used for the performance of the other tasks of the project. The feedstocks utilized in the study were a wide range middle distillate (149-385°C) and the remaining gas oil (385°C+) from the whole syncrude. The middle distillate boiling range was selected to represent a probable distillate of the future. The gas oil IBP is as a result of the distillate range. The gas oil also includes the 524°C+ residuum material described in the assay.

The two feedstocks were prepared for the two modes of catalytic processing in the project and therefore there are two separate sets of results; one for hydrotreating synthetic crude middle distillate and one for hydrocracking synthetic crude gas oil. Both the hydrotreating and hydrocracking work was focused on three basic areas:

- Screening of commercial catalysts
- Catalyst selection and condition optimization
- Production run and evaluation

4.1 Crude Oil Evaluation

1) Crude Oil

The synthetic crude oil was a 31.5°API gravity crude with a total sulfur content of 0.15 wt%. The crude consisted of 2.89 vol% light ends (C_3 and C_4's by GCD) and had the following yields:

	ASTM D2892 Yield Vol%
IBP-15°C	2.33
Straight Run Naphtha (15-149°C)	15.70
Atmospheric Distillate (149-385°C)	60.42
Gas Oil (385-524°C)	20.39
Residuum (524°C+)	1.16

2) Naphtha (15-149°C)

The sulphur and nitrogen content were 0.02 wt% and 1 ppm (wt/wt) respectively. The light ends (IBP-15°C) were removed before further distillation and were found to be 2.33 vol% of the crude. Compositional data on this naphtha called "straight run" is included in this portion of the report. For comparison and convenience the HT- and HC-derived naphthas are also shown.

3) Middle Distillate (149-385°C)

The yield of atmospheric middle distillate from the crude was 60.42 vol% with a 28.6°API gravity. The distillate fraction had a sulphur content of 0.10 wt% and a nitrogen content of 167 ppm wt/wt. The cetane number 30 indicated poor diesel fuel ignition and combustion properties. The pour point (below -51°C) and a cloud point (-33°C) showed good low temperature operability. This material was used as feed in the hydrotreating portion of the contract.

4) Gas Oil (385°C+)

The yield of gas oil from the synthetic crude was 20.39 vol% with a 16.6°API gravity. This material was used as feed in the hydrocracking portion of the project.

Attachments to this synthetic crude evaluation are contained in Section 7 of this report. They include Table 4.1.1, the Crude & Product Analysis Summary; Figure 4.1.1, the Crude TBP Curve and Tables 4.1.2 - 4.1.5 which are supplementary information for the evaluation.

Table 4.1.2 is a summary of a reformer model prediction for the straight run naphtha. It shows the predicted yields of typical platformer products for various levels of production severity. Table 4.1.3 is a detailed mass spectroscopy report on the middle distillate and Table 4.1.4 is the same type of report for the gas oil fraction. Table 4.1.5 reports the gas chromatographic distillation on the whole crude.

4.2 Hydrotreating Synthetic Crude Middle Distillate

4.2.1 Catalyst Screening

A field of eight commercially available catalysts was selected and screened for the most effective middle distillate hydrotreatment performance. The catalysts selected were:

American Cyanamid HDN30
Harshaw NI4301E
Katalco NM506
Katalco NT550
Ketjen Armak KF8400
Shell S324
Shell S454
Gulf Catalyst A

The catalyst properties are listed in Table 4.2.1 and it can be seen that these catalysts are all extrudates of consistent size with varying quantities of metals and surface properties. The Gulf Catalyst A properties are not listed but are similar to those described in U.S. Patent No. 4,427,534 and is a catalyst manufactured under license by Harshaw. The intent of this selection was to observe the directional effects of various catalyst properties in syncrude middle distillate hydrogenation. The field includes 4 nickel/molybdenum catalysts (HDN30, KF840Q, NM506, S324) and 4 nickel/tungsten catalysts (NT550, NI4301E, S454, Gulf A); a variety of surface areas from 134 to 230 m^2/g; 2 catalysts with silica-alumina supports and a wide range of metal contents.

The P-16 pilot plant hydrotreater was used to treat the middle distillate feedstock over a range of three treatment temperatures (340, 385 and 420°C) at a pressure of 13.8 MPag and a 1.0 LHSV for each catalyst. The pilot hydrotreater was a single, fixed bed unit containing approximately 170 mL catalyst and operating as a non-recycle or once through hydrogen system. The hydrotreater was operated consistently for each of the catalysts evaluated from loading, sulphiding and de-edging through to run lineout, production and product sampling.

The collected individual run products were analyzed for sulphur, nitrogen, and boiling point distribution as well as molecular characterization by mass spectroscopy.

The collected data for each catalyst is presented in Tables 4.2.2 to 4.2.9. From these tables it can be seen that a wide range of performance exists over the field of eight catalysts.

In order to discern trends in the results, the following plots were made:

Figure 4.2.1 Aromatic Conversion vs Reaction WABT*
Figure 4.2.2 Total Liquid Product M.W. vs Reaction WABT
Figure 4.2.3 Wt% Ni on Catalyst vs Aromatic Conversion
Figure 4.2.4 Catalyst Surface Area vs Aromatic Conversion

* WABT = weight average (catalyst) bed temperature

The objective of the middle distillate hydrotreating program was the achievement of maximum aromatic conversion while minimizing the cracking of feed into naphtha. The catalyst screened which offered the best combination of these was to be chosen for optimization. The rationale for these criteria were that the severity and hence cost of this hydrotreating configuration would be sacrificed if the operation produced excessive amounts of naphtha. On the other hand, treatment severity is necessary to upgrade syncrude middle distillate to an acceptable quality. The penalty imposed by producing naphtha from the hydrotreating operation is the cost of processing itself and the loss of distillate yield as well as the need for further processing of the naphtha to make saleable gasoline. As such, naphtha would be a loss to the distillate refining operation. Processing severity (temperature, pressure, LHSV) in upgrading syncrude middle distillate is required for the saturation of its large aromatic content. The resultant production of cycloparaffins is related to the ignition and combustion property improvement in synfuel product.

Other treatment reactions observed in the catalyst screening runs were hydrodesulfurization and hydrodenitrogenation, but these were essentially complete at the mildest severity run conditions. Therefore, most of the observations made were directed at understanding aromatic conversion in the middle distillate.

From the literature (3), an approximate hydroprocessing hierarchy of severity for major reactions groups (not considering overlaps) is suggested. In order of increasing severity of processing:

Heteroatom removal
- sulphur and nitrogen removal (HDS, HDN)
  e.g. thiophenes: single ring followed by di, tri etc.

Aromatic Saturation
- polynuclear aromatics first and increasing severity to monoaromatics

Hydrocracking
- alkyl side chains
- scission of multi-ring cycloparaffins
- paraffins
- scission of multi-ring aromatics
- ring opening

To achieve good product quality, the objective becomes optimization of the aromatic saturation reactions and minimizing hydrocracking reactions. This is to justify the selection of operating conditions for the catalyst screening exercise. The pressure and space velocity were fixed for relative comparison of catalysts at a minimum number of conditions. Temperature was therefore varied to produce mild, moderate and severe operating conditions in expectation of observing the associated molecular changes by mass spectroscopy.

From Figure 4.2.1 it can be seen that the catalysts provide aromatic conversion in a non-linear response to reactor temperature. Over the range of operating temperatures, there appears to be an optimum aromatic conversion point reached at a mid range temperature (approx. 385°C). This is true except for the Harshaw and Gulf catalysts which increase conversion with increased temperature. Figure 4.2.2 shows the effect of reactor temperature on the molecular weight of the treated product and the relative activity of each catalyst can be seen. It is evident that the Harshaw and Gulf catalysts are significantly different in performance characteristics stemming from their supports being silica-alumina rather than just alumina like the supports of the rest of the catalysts. The properties of the Harshaw and Gulf catalysts otherwise are not significantly different in metals composition or surface areas compared to the Katalco catalysts in particular. The higher nickel contents of Ni4301E and Gulf A do not seem to be in proportion to the effect on product molecular weight, hence the support

must supply the cracking function. This is understandable in light of the use of silica-alumina for FCCU catalysts prior to the introduction of zeolites (3) and their use in other hydrocracking operations (7). The excessive cracking produced by silica-alumina catalysts in the screening runs makes them largely unsuitable for upgrading syncrude middle distillate.

The apparent optimum in aromatics conversion for the alumina supported catalysts would appear to occur at roughly the same temperature for all of this group. This is due to an equilibrium in the aromatic saturation reaction where the conversion declines with the equilibrium shift associated with temperatures beyond 385°C. The temperature response of the group shows that different sensitivities exist as well. Shell S454 shows the lowest sensitivity in aromatic conversion in response to reaction temperature. A review of Table 4.2.1 revealed that S454 had a small amount of silica in its support composition and this may have had the effect of a mild hydrocracking catalyst which shows minor improvements in aromatic saturation at the high temperature. S454 also gave relatively higher naphtha makes. The most sensitive catalysts look to be HDN30 and Shell S324. It was of interest to examine if there were any trends related to catlayst properties.

Figure 4.2.3 shows directionally that the higher the concentration of nickel promoter in the catalyst the higher the aromatic saturation at the optimum conversion point. Figure 4.2.4 shows catalyst surface area trends with aromatic saturation. Plots of other catalyst properties vs. aromatic conversion were looked at, but trends were not evident.

However, isolating one catalyst property and relating it to a performance criteria is somewhat simplistic and further work in catalyst property development/optimization would be necessary to distinguish real trends. The relatively poorer performance of a particular catalyst in aromatic saturation could be because the intended use was different from that of our application. Each catalyst in this set has been specified for a particular hydrotreating purpose.

Looking in further detail, the comparison of mass spectroscopy results are shown in Tables 4.2.10-4.2.17 for each catalyst screened. These results are for fractionated products, where the

naphtha has been removed. The naphtha yields are given in the run summary tables referred to earlier. It should be noted that the Paraffin/Naphthene/Aromatic (PNA) distribution in the naphtha is slightly lower in aromatic content and relatively higher in paraffins than in the corresponding middle distillate. Removing the naphtha from the run products concentrated the aromatics but gave a truer picture of production middle distillate.

The run to run change of molecular distribution for each catalyst is compared in the mass spectroscopy tables (4.2.10-4.2.17) and looking at the NM506 runs in particular (Table 4.2.12), one can observe the molecular changes associated with severity. Comparison of the feed (Table 4.1.3) to the lowest severity run shows a significant decline in mono- and diaromatics, as well as thiophenes. The low severity run has accomplished heteroatom removal and aromatic saturation.

The mid-range temperature in the NM506 catalyst case indicates a maximum for aromatic conversion and the point where the onset of hydrocracking reactions become apparent. The product compositional analyses from the higher temperature run showed an increase in aromatics in response to the reversal of the aromatic saturation equilibrium and a higher level of hydrocracking was demonstrated with an increase of naphtha make over the 385°C run. The product compositional analyses (420°C) also showed slightly lower levels of cycloparaffins present and this was further evidence that the aromatic saturation equilibrium had shifted.

The above observations, however, are valid only for the alumina supported catalysts and for these catalysts the Ni content would seem to be related to the level of aromatic conversion achieved (Figure 4.2.3).

The silica-alumina supports of the Harshaw and Gulf A catalysts appeared to promote strong cracking performance even at moderate severity. The naphtha make was large, exceeding 50% at high severity. However, the aromatic conversion does not seem limited. This was particularly true of Gulf A catalyst, which was not even screened at the highest severity because the naphtha make was already 20 vol% at mid-severity. The silica-alumina support cracked substituted and multi-ring aromatics to the point where ring opening was observed and the large naphtha make was the result of this support's high level of cracking activity. But in light of

the given criteria for selection of the hydrogenation catalyst, the Harshaw and Gulf catalysts produce too much naphtha and are very sensitive to temperature making them more difficult to operate. Therefore the alumina supported catalysts appear to be preferable for the middle distillate upgrading operation.

Amongst the alumina-supported catalysts, the highest aromatic conversion was achieved using Katalco NM506. The corresponding naphtha make was minimal, therefore satisfying the selection criteria.

Coincidently NM506 had the highest Ni content, and highest surface area of the alumina group as well as good bulk density. The high surface area and bulk density characteristics are desirable from the standpoint of achievable reactivity per unit volume of catalyst bed.

A recommendation to use Katalco NM506 in the further treatment studies of the project was made to the Scientific Authority. With catalyst choice agreement, the program moved on to the optimization of reaction conditions.

4.2.2 Hydrotreating Condition Optimization

The chosen catalyst, Katalco NM506, was prepared in the P-12 unit hydrotreater for the condition optimization work of Task No. 3. The agreed plan of conditions were three temperatures (335, 360, 385°C) at three different pressures (10.3, 13.8, 17.2 MPag) in order to map out the directional effects of these parameters. Two supplementary runs were done at 410°C at 13.8 and 17.2 MPag for added definition. The LHSV utilized remained fixed at 1.0, to reflect a minimum commercial treatment rate and to provide the most head room for distillate upgrading. Also this would help minimize mass transfer effects in the microreactor so that molecular changes could be tracked more reliably. Higher space velocities, especially at lower pressures, cause changes in trickle-bed flow regimes (particularly gas flows) in microreactors (4).

The catalyst bed length in the P-12 unit was shorter than in the P-16 arrangement and had the effect of reducing aromatic conversion at identical operating conditions. The well known Mears' design criteria (5) gave an indication of the effect of bed length on conversion:

$$\frac{L}{d_s} > \frac{20}{Bo} \ln \frac{C_o}{C_f}$$

where L = reactor bed length
d_s = catalyst particle size, spherical equivalent
Bo = Bodenstein number of the fluid system (at T & P)
C_o = reactant concentration in the feed
C_f = reactant concentration in the product

The L/d_s ratio was approximately 400 on the P-16 unit versus 300 on the P-12 unit. The catalyst bed of P-12 was operated at the maximum loading for this unit, but resulted in somewhat lower aromatic conversion than those observed in the P-16 screening runs. It is thought that the difference in catalyst bed length was the principal reason for the lower conversions observed. However, P-12 has been used quite reliably in the past to simulate commercial treatments and, with its recycle gas capability, is similar to a full scale unit. The aromatic conversion at 385°C and 13.8 MPa was 12% lower on P-12 compared to P-16. This reduction is predicted by the Mear's relationship almost exactly. But P-12 yields should be considered commercially feasible, and the product to be representative.

The results of the optimization runs are summarized in Tables 4.2.18, 4.2.19 and 4.2.20 and plotted in Figure 4.2.5.

Directionally, the results support the treatment trends expected with the best aromatic conversion occurring at the highest pressure. The temperature effect on aromatic conversion repeated the pattern observed in the screening runs and the optimum conversion for each pressure occurred close to 385°C reactor temperature. The aromatic conversion optimums for each pressure are plotted in Figure 4.2.6 and it can be seen that diminishing gains are made with increasing pressure. The aromatic conversion gain between 10.3 and 13.8 MPag is more significant than the conversion gain going from 13.8 to 17.2 MPag. The extrapolation of this curve to 20.7 MPag indicates that a reduction in total aromatics from 6% to 4% would occur and predicts that conversion gains will diminish with increasing pressure. The offsetting benefit in increasing pressure (at 385°C), however, is the increase in middle distillate yield as the following comparison shows:

Pressure	17.2 MPag	13.8 MPag	10.3 MPag
H_2 Uptake, m^3/m^3	150	138	109
Naphtha Make, vol%	2.5	3.0	3.1
Yield, vol% TLP	105.2	104.8	103.9

The uptake of hydrogen and total liquid yield increased with increasing pressure. Naphtha yield did seem to decrease with pressure. Reviewing the apparent product changes in response to pressure in more detail, the mass spectroscopy results of Table 4.2.21 show the effect of reaction pressure on product composition for the 385°C runs.

Apart from the aromatic saturation differences, the molecular changes in response to pressure on the NM506 catalyst are not dramatic. However, it should be noted that the level of paraffin content (compared to the feed) actually increases at low pressure and declines as the pressure increases. A possible explanation for this is that the cracking of alkyl side chains from substituted cycloparaffins or aromatics is adding to the population of paraffins at low pressure. The increase in pressure and subsequent lower paraffin contents is therefore indicative of the suppression of cracking. The change in aromatic size distribution is also interesting. The effect of pressure on larger aromatics (tri+) is almost negligible, the effect on diaromatics is marginally better and the effect on monoaromatics is the most significant. This would suggest that the most pronounced effects of pressure in treating this syncrude middle distillate is the improvement of monoaromatic saturation (for this catalyst). Similar observations can be made on the effect of pressure at other temperatures with the exception of runs 2020 and 2029 where the lower pressure (13.8 MPag) product had lower aromatics than the high pressure (17.2 MPag) product. Figure 4.2.6 shows conversion isotherms and the 335°C runs indicate a conversion optimum at 14.5 MPag which may explain the crossover of the isotropic curves in Figure 4.2.5. The general shape of the aromatic conversion curves in Figure 4.2.5 are otherwise concentric and product from run 2029 should have had lower aromatics to fit the conversion pattern that is otherwise evident. The operating pressure affects the aromatic saturation optimums in the 360 and 385°C runs as well. The respective pressure optimums appear to be 15.8 and 19.5 MPag (by extrapolation).

Touching briefly on the commercial aspects, the high pressure reactor required for the highest aromatic saturation achieved in this pilot study (17.2 MPag) would be expensive from both capital and operating points of view. However, the ignition and combustion property improvement needed to produce quality fuels from syncrude middle distillate does require aggressive treatment. In the case of diesel fuel product, the middle distillate feedstock has a cetane number of 34 to 36. The current CGSB product specifications require a minimum of 40 cetane number in diesel fuel. The high aromatic conversion achieved at the 17.2 MPag operating pressure actually produces a diesel fuel which exceeds the product minimum specification by 6 cetane numbers. Therefore, to offset the higher capital and operating costs of the higher pressure process, the space velocity could be increased to where the product reached 40 cetane. This would reduce the size, hence cost, of the reactor system for a given production level. As mentioned before, relative gains in yield are also enjoyed at high pressure and would also compliment the higher pressure operation. The diminishing returns observed, of going beyond 17.2 MPag, suggest that this pressure was possibly a practical limit for the process scheme.

In light of the foregoing discussion, it was recommended to the Scientific Authority that the high pressure (17.2 MPag) and the observed optimum temperature (385°C) conditions be adopted for the production run.

4.2.3 Hydrotreating Production Run

The raw syncrude had 60.4 vol% of a full range middle distillate fraction. The nominal cut point was 149-385°C. As the syncrude had a low wax content, higher end point diesel fuel could be produced. There were over 200 litres of middle distillate available for processing and the production runs occurred in April and May, 1984. Approximately 40 litres of product were produced, over a four week period.

Conditions chosen for the production run were established to maximize aromatic conversion, and minimize cracking. The pressure used was 17.2 MPag and the run was performed on the P-12 pilot plant unit. Yield data from the production run 2035 is shown in Table 4.2.22. TLP volume yield was 106.1 vol%, with a hydrogen consumption of 158 m^3/m^3. The yield of 149°C-naphtha was 7.7

volume percent. All the TLP samples were composited prior to fractional distillation. The production run product was quite similar to that produced in the condition optimization runs, and these are also given in Table 4.2.22 for comparison.

Based on the quality of middle distillate feed from 1 m^3 of syncrude, the hydrotreating mode would produce the following ultimate yields:

		Product
Feeds of 604.2 litres (149-385°C)	IBP-15°C	2.6 litres
+ 95.9 m^3 H_2	15-149°C	47.1 litres
	149-385°C	597.8 litres
Total		647.5 litres

It is evident from these production runs that with suitable conditions, it is not only possible to make a high cetane number diesel fuel from this feedstock, but also to produce high yields. The diesel product was 98.9 vol% of the middle distillate feed.

4.2.4 Hydrotreated Product Evaluations

The liquid products from the middle distillate hydrotreating production run consisted of:

	Litres	Vol%
IBP-15°C	0.2	0.4
Naphtha 15-149°C	3.1	7.3
Middle Distillate 149-385°C	39.2	91.3
	42.5	100.0

Product Quality

a) Naphtha

The total volume of hydrotreater produced naphtha was relatively small. On the basis of 100 bbls of syncrude feed, there would only be 7.8 bbls of the HT naphtha produced. The volumes of this naphtha, along with the HC (hydrocracker) naphtha and SR (straight run naphtha) were combined in our crude assay report. It was felt that commercially it would not be practical to segregate the HT naphtha, rather it would be combined with the SR and HC naphtha for further processing. However the separate composition is shown below:

HT Naphtha Composition

nominal boiling point		15-149°C
S ppm		0
N ppm		0
°API		56.7
Specific Gravity		0.752
Average M.W.		102.8
V.P. kPa @ 40°C blending F1/F2		49.2
Vol% C_6^+ fraction	P	27.1
	N	67.7
	A	5.2
Vol% of total naphtha		
	iC_4	0.7
	nC_4	3.0
	iC_5	2.4
	nC_5	3.5
	Cyclopentane	0.4
	C_6^+	90.0

The HT naphtha has more naphthenes than conventional naphtha, and would produce a highly aromatic reformate upon catalytic reforming. The mass spectroscopy compositional is given in Table 4.2.23. More detail on gasoline quality is given in the naphtha disposition section (4.4).

Jet B Fuel

This was blended from 20% HT 149-300°C middle distillate and 80% HT naphtha. The naphtha was depentanized. Data is shown in Table 4.3.9.

The Jet B fuel production would usually need some C_5's back blended to meet the minimum RVP specification. As it was prepared, the RVP was 5.9 kPa. Otherwise, the product appears acceptable. It is somewhat heavier than conventional Jet B fuel due to the high content of cycloparaffins. The luminometer number for this fuel was high compared to the specification minimum and suggested that good combustion properties are characteristic of this Jet B blend.

Jet A Fuel

The hydrotreated middle distillate from the production run was fractionated to the CGSB maximum end point of 300°C. The resultant fuel had a specific gravity that was at the CGSB maximum as well.

The fuel specification test results are shown in Table 4.3.10 along with the hydrocracking production run product Jet A fuel. It can be seen that the Jet A fuel produced in the hydrotreating operation meets or exceeds all specification tests. The luminometer number is good at 55 and suggests that combustion of this fuel should be good as well. The conductivity result would be altered with an appropriate additive to fall in the specified range.

In all respects this Jet A fuel is quite acceptable for retailing by current standards. The only close point of concern is perhaps the high viscosity result at -20°C, even though it is below the maximum specified. The viscosity is the result of the high density and naphthene content. It could be offset by blending or changing the selected distillation end point.

Diesel Fuel

In order to produce jet and diesel fuels from the same middle distillate, an assumption on volume ratios was made. For the purpose of this project, a sales volume ratio of 25/4/1 Diesel/ Jet A/Jet B was assumed so as to form a basis for re-distribution of the 300-385°C fraction left over in the preparation of the jet fuels. This resulted in the hydrotreated diesel fuel being a composition of 94 vol% full range middle distillate product and 6 vol% 300-385°C bottoms from the jet fuel fractionation. The CGSB specification test results are given in Table 4.3.11 (CAN2-3.6-M83 Type B).

As can be seen in this table, the hydrotreated diesel fuel had a good cetane number of 45 (by engine test) and a pour point which satisfies type A and AA specifications. However, because the prepared blend was a wide range diesel, type A and AA 90% distillation criteria are not met. Nonetheless, the diesel fuel would make satisfactory low temperature product. As with the jet fuel, the viscosity of this diesel product is high in the range for type B and as such may pose problems in some end uses.

Cold weather use of this material may require changing the end point and/or blending with lower viscosity product. As is, the diesel fuel would probably satisfy market requirements down to -30°C.

In comparison to conventional diesel fuel, the hydrotreated syncrude middle distillate is much higher in naphthene content due to the paraffin deficiency in the raw distillate. Cetane improvement in syncrude distillate is entirely due to aromatic saturation to naphthenes which contribute relatively less to cetane rating than paraffins. Consequently with aromatics reduced to less than 6 wt% in the syncrude (hydrotreated) diesel little headroom for cetane improvement (beyond the measured 45) is possible without conversion of cycloparaffins to paraffins.

As was the case with the Jet A fuel, an anti static additive would be needed to improve the conductivity of the fuel.

The diesel fuel had a specific gravity of 0.8408 and the D86 distillation is shown below:

Vol % Distilled	Boiling Point, °C
IBP	179
10	205
30	237
50	267
70	297
90	328
FBP	348

4.3 Hydrocracking Synthetic Crude Gas Oil

4.3.1 Introduction

The 385°C+ fraction of the syncrude is referred to as gas oil, and this fraction was dealt with separately. Its volume was 21.5% of the total syncrude, so it is the second most significant fraction or portion of the crude. Indeed, the processing of this stream was perhaps the most interesting and significant part of the entire project.

While the properties of this material are covered elsewhere in this report, some discussion is warranted, as the gas oil is somewhat unique compared to conventional gas oils. This material was similar to the 385°C+ gas oil that is listed in the crude oil evaluation, but the small percentage of bottoms that appeared in the assay was left in the gas oil for the processing section of the work.

Properties of syncrude gas oil

°API	16.7
Specific Gravity	0.9549
Sulfur	0.4 wt%
Nitrogen	1510 ppm
Metals	.05 ppm Ni, .07 ppm V

	LRMS, syncrude gas oil (Mass %)
Naphthenes	36.0
Paraffins	4.8
Aromatics	52.1
Aromatic Sulfur	2.9
Polar Compounds	3.6
Unidentified	0.7
	100.0

It should be noted that this is a very heavy, aromatic type gas oil. It is more aromatic than most fluid catalytic cracking unit feedstocks, and for this project it was decided to evaluate it as a hydrocracking feedstock only. Approximately 100 litres of the gas oil material was accumulated, and it formed the basis of our evaluation as outlined in the work statement.

4.3.2 Catalyst Screening Runs on a Microreactor Unit

Two gas oil hydrocracking catalysts were chosen, and the runs were carried out in the P-16 unit located in the pilot plant.

The catalysts chosen for the screening runs were Gulf Catalyst "A" and Katalco NT-550.

- Gulf Catalyst A. This is a high activity Gulf catalyst. It is noted that it is the most active hydrocracking/saturation catalyst produced by Gulf and it has a long history of successful commercial operation. It is a Ni/W catalyst with a silica-alumina base and is fluorided.

- Katalco NT-550. This also is a commercially available catalyst. It is referred to as a nickel-tungsten hydrotreating catalyst, and has an alumina base. It is non-fluorided.

With the program set out with two catalyst choice options, a catalyst was chosen from each of the two catalyst support types available in the selected field. The alumina support was decided upon in order to observe the comparative effects of this type vs. a silica-alumina support in gas oil hydrocracking. Also the selectivity of alumina to produce middle distillate vs. naphtha was to be tested. The NT-550 catalyst was chosen from the alumina group based on known past performance as an HDN, HDS and HT catalyst. The silica-alumina (Gulf Catalyst A) was chosen based on long and good service in hydrocracking operations representing a mature, high activity catalyst for gas oil hydrocracking. The Gulf Catalyst A activity was a concern, however, and the selectivity for middle distillate was in question.

The properties of both these catalysts are listed in Table 4.2.1.

The screening runs for each were done over a seven-day period consisting of 1 day set up and 1 day presulfiding and de-edging.

The nominal 385°C+ cut gas oil was used. In each case the LHSV was 0.5. Both catalysts showed good hydrodesulfurization and hydrodenitrogenation properties.

The data collected includes yields, light ends, TLP distillation, molecular weight, sulphur, nitrogen, and compositional mass spec. The data is given in Tables 4.3.1- 4.3.2. Figure 4.3.1 is a plot of gas oil conversion versus reactor temperature and illustrates the relative activities of the cracking catalyst.

The above two catalysts were chosen to investigate the product yield and qualities from gas oil hydrocracking. Factors considered in this selection were the base material of the catalyst, e.g. silica-alumina with fluoride promotion vs. alumina, and catalyst selectivities for middle distillate production.

After reviewing preliminary data, priorities in the selection of a catalyst included activity, selectivity, and catalyst life. From this standpoint, the catalyst with the lowest start of run operating temperature, for a given conversion, is more active and offers the most headroom for activity loss corrections using temperature. The operating headroom offers potentially longer useful life. The selectivity of a catalyst in hydrocracking operation can be seen as the successful cracking of larger molecules while avoiding production of light ends and coke.

If activity/selectivity criteria is used to rate the performance of the two tested catalysts, then the following comparison can be made. Figure 4.3.1 shows selected data taken at constant LHSV (0.5), pressure (20.7 MPag), and gas rate (890 L/L). A comparison of performance at constant conversion illustrates activity, for example:

	60% Conversion	
	Gulf A	NT-550
Est. Temp. °C	391	409
Cracked Light Ends, %	4.0	4.0
TLP Aromatics, %	4.5	3.7

From a processing point of view, it was felt that the 18°C start of run temperature difference is significant due to the potential for longer runs in the fixed bed.

For any given temperature within the range tested, the conversion was highest with Gulf Catalyst A. The selectivity criteria can be similarly reviewed by comparison of the light ends production. In the example case given here, there appears to be little difference in produced light ends as shown by Figure 4.3.1B. This suggests that these catalysts probably have similar selectivities. One difference, however, is the temperature at which each catalyst produces these light ends. The higher the operating temperature, the higher the risk of coking reactions leading to catalyst deactivation.

It is clear from this comparison that Gulf Catalyst A offered the lowest start of run temperature for a given cracking level, good aromatic conversion and light end losses comparable with NT 550. It should have the longest run life for a given conversion level because of its relatively higher initial activity. For the criteria mentioned, it was therefore recommended that the hydrocracking production run proceed using Gulf A as the catalyst.

4.3.3 Hydrocracking Optimization Runs

Having selected the most effective of the two catalysts studied, a program to investigate the effects of the process variables, pressure and temperature on product yield and quality was carried out.

The conditions chosen for evaluation were:

Temperature, °C	Pressures, MPag
360	13.8
385	17.2
400	20.7

For each pressure, runs were done at 3 temperatures, so there were a total of 9 runs. In addition, two extra temperature runs were done at the 20.7 MPag level - 410 and 420°C. The run results are contained in Tables 4.3.3-4.3.6. and are shown graphically in Figure 4.3.3.

The May 15th progress meeting allowed interactive discussion on the hydrocracking results and the Scientific Authority agreed to choosing conditions which maximized middle distillate yield. The

production run conditions were thereby set as 405°C and 20.7 MPag. The directional indication of Gulf Catalyst A performance was seen to be better at higher operating pressures (for a given temperature) and the highest middle distillate yield occurred at 405°C. This was then chosen as the basis for the production run conditions. Beyond this temperature excessive naphtha and light ends were produced. The supplementary runs were performed to define this optimum for 20.7 MPag (the most active pressure).

The product from the run representing the chosen optimum conditions had a conversion level of greater than 80% to lighter material and a volume yield better than 110%. The aromatics content of the 385°C-product as measured by mass spectroscopy was less than 6% (this included some low aromatic naphtha, approx. 15 vol%).

The mass spectrometric analysis of the middle distillate fraction compares well with that of the production run product from hydrotreating. If this is an indication, then we expected a cetane number in the 40's from the hydrocracker distillate.

The effect of aromatic saturation was:

a) Wt% aromatics in 385°C+ reactor effluent (unconverted gas oil)

Pressure	Temperature, °C		
MPag	360	385	400/05
20.7	44	23	5
17.2	48	33	26
13.8	51	45	24

The lowest aromatic product occurred at the highest pressure, and 400°C.

Denitrification of the TLP (total liquid product) was:

b) Wt ppm nitrogen in TLP

Pressure	Temperature, °C		
MPag	360	385	400/05
20.7	12	7	1
17.2	52	19	2
13.8	31	19	5

Again, the lowest nitrogen product occurred at the highest shown temperature. Pressure seemed not so critical.

Desulfurization was as follows:

c) Wt ppm sulfur in TLP (H_2S excluded)

Pressure	Temperature, °C		
MPag	360	385	400/05
20.7	90	23	3
17.2	90	77	18
13.8	118	96	35

High temperature and pressure favoured desulfurization.

4.3.4 Hydrocracking Production Run

Approximately 30 litres of the raw syncrude gas oil feed was available for the final "production run". This run occurred during a 3-week period in June, 1984. All the TLP was composited for fractionation prior to component analysis. The P-16 unit was used for the entire production run.

The run produced 4.3 L naphtha (149°C-), 19.8 L of middle distillate (149-385°C) and 8.0 L of unconverted gas oil (385°C+). The yield of total liquid product on the feed was approximately 110 vol% with a hydrogen consumption of 225 m^3/m^3 (1265 scfb). Table 4.3.7 shows the production run data with the Task #4 data at comparable conditions. It can be seen that the production run TLP showed slightly less gravity gain than the optimization run. This is considered acceptable reproducibility. The middle distillate from the production run had a measured cetane number of 44.5 which indicated that Gulf Catalyst A as a hydrocracking catalyst performed well for aromatic conversion as well as hydrocracking. The blending evaluation in the next section elaborates on the fuel quality of the production run products.

4.3.5 Hydrocracked Product Evaluations

The liquid products from the hydrocracking production run consisted of:

	Litres	Vol%
Naphtha 149°C-	4.3	13.4
Middle Distillate 149-385°C	19.8	61.7
Gas Oil 385°C+	8.0	24.9
	32.1	100.0

Product Quality

a) Naphtha

Before addressing quality and properties of motor gasolines and aviation Jet fuel, it is important to be aware of the total potential volume of this naphtha component and therefore its relevance to potential markets. Refer to Naphtha Disposition section 4.4 for further details.

As only 4 litres were produced, tests were limited. In the overall scheme, it was shown that the gas oil feed to the hydrocracker was 21.5% of the original synthetic crude. Thus for 100% of syncrude charge, we would get 21.5 x 13.4% x 1.10 = 2.6% of hydrocracker naphtha. So it would not be practical to segregate this naphtha, rather it would be recombined with the straight run and hydrotreated naphtha for further processing.

Hydrocracker Naphtha Composition

The mass spectroscopy analysis is attached in Table 4.3.8. The C_5 and lighter portion was 8.6 vol. percent of the total HC Naphtha.

Hydrocracker Naphtha (Sample No. 6072)

Nominal Cut Point		15-149°C
S, ppm		0
N, ppm		0
°API		58.2
Specific Gravity		0.746
Average MW		100.3
Calculated V.P. kPa @ 40°C		31.9
Blending F1/F2		61/60.5
Vol.% on C_6+ Fraction	P	29.7
	N	66.7
	A	3.6
Vol.% on C_5- Fraction	iC_4	0.43
	iC_4	1.29
	iC_5	3.00
	nC_5	3.23
	Cyclopentane	0.61
C_5 and lower;	Calculated F1 Octane	81.0
	Calculated F2 Octane	79.4

The hydrocracker naphtha has more naphthenes than conventional naphtha, and less paraffins and aromatics. For catalytic reformer feedstock, the HC naphtha feed has a potential to produce a highly aromatic product, as the naphthenes would be dehydrogenated to aromatics.

Jet B Fuel

This was blended from 20% HC Jet A and 80% HC naphtha. The naphtha was depentanized. Results of the CGSB specification tests are contained in Table 4.3.9. The high luminometer No. (172.2) was an indication of the low aromatic content.

Jet B production would need some C_5's back blended to meet the minimum R.V.P. Otherwise the product appears acceptable. It is somewhat heavier than conventional Jet B due to the high content of cycloparaffins.

b) Hydrocracker Middle Distillate Product Quality

Jet A Fuel

In order to meet the maximum density specification of 0.839 g/cc, the middle distillate blended to Jet A was refractionated to produce a 281°C E.P. material. Approximately 44% of the 385°C E.P. middle distillate was rejected to produce the Jet A fuel. Tests showed that this material met the existing Jet A specifications. The borderline points were viscosities, WSIM and density. Refer to Table 4.3.10 for the complete listing. Conductivity was very low, as is usually the case with hydrotreated product. An antistatic additive would be used commercially to bring it on specification.

Diesel Fuel

The full range middle distillate easily met the CAN2-3.6-M83 Type B diesel specification. The cetane number was 46 by engine test. The pour point met the Type A and AA specifications, but the cloud point and 90% B.P. did not. Nonetheless, their full range material would make an acceptable low pour diesel for temperatures down to -27°C. The full tests are reported in Table 4.3.11.

As was the case with the Jet A fuel, an antistatic additive would be recommended to improve the conductivity of the fuel. The diesel fuel relative to commercial diesels was highly naphthenic.

The production run diesel sample had a specific gravity of 0.8608 (32.9°API). The ASTM D86 distillation was:

IBP	178°C
10%	210°C
30%	225°C
50%	296°C
70%	327°C
90°C	352°C
FBP	360°C

c) Unconverted Gas Oil

Since the unconverted gas oil was not considered a product, the project did not allow for its testing. Its volume was 27.4% of the raw gas oil feed. It is a low wax, low aromatic stream and

has a high cycloparaffin content. It would make a good HC recycle feed or FCCU feed. It has low sulfur and nitrogen as it has been through the hydrocracker conditions.

Test results on the unconverted gas oil were:

Specific Gravity	0.8786
°API	29.5
S	44 ppm
Nitrogen	5 ppm
Pour Point	+12°C
Viscosity @ 40°C	46.2 cSt
@ 100°C	6.5 cSt

d) Mass Spectrometer Data

For naphtha and Jet B fuel, see Table 4.3.8.

For Diesel fuel and Jet A fuel, see Table 4.3.7.

For 385°C+ fraction, see Table 4.3.7.

4.4 Naphtha Disposition

In this project three types of naphtha were produced:
1) SR Naphtha from syncrude fractionation
2) HT Naphtha from middle distillate hydrogenation
3) HC Naphtha from gas oil hydrocracking

The properties of these naphthas were listed from lab information and our monthly reports.

Due to the small amount of HT and HC naphthas, they were lumped together with the SR material and on a depentanized basis the combined properties were calculated. The combined PONA of 44.0, 48.6 and 7.4 shown on Page 51 is quite close to that of conventional naphtha.

The C_5 octanes were also calculated and it was shown that the reformate would have to be 95 RON to meet the C_5 and reformate blend of regular unleaded $\frac{R+M}{2}$ of 87.

SUMMARY OF NAPHTHA PROPERTIES FROM SYNCRUDE PROCESSING

This list represents properties of the 3 naphthas produced in this program.

	SR Fract.	HT Production Run No. 2159	HC Production Run No. 6072
Nominal cut point	15-149°C	15-149°C	15-149°C
Vol % on crude	15.7	--	--
Sulphur, wt%	0.02	nil	nil
Nitrogen, ppm wt	1	nil	nil
D86 - IBP	39		
10	57		
30	72	N/A	N/A
50	88		
70	106		
90	128		
FBP	159		
RON/MON	65.0/62.0	61.6/61.0*	61.0/60.5*
Light Ends, vol%			
iC_4	0.15	0.7	0.43
nC_4	2.94	3.0	1.29
iC_5	7.55	2.4	3.00
nC_5	11.8	3.5	3.23
Cyclo C_5	2.2	0.4	0.61
C_6^+	75.4	90.0	91.4
C_6^+ Portion - P	54.5	27.1	29.7
- N	36.2	67.1	66.7
- A	9.3	5.2	3.6
Calc. Properties,			
Density	0.712	0.752	0.746
Avg. Mole. Wt.	91.5	102.8	100.3
VP kPag @ 40°C	55.6	49.2	31.9

*Octane Numbers are values determined from engine tests on mixtures with SR naphtha.
N/A = not available.

The following calculation shows the composition of the combined naphthas on a depentanized basis.

Naphtha Production Basis: 100 cubic metres of crude charge

	Blended from Integrated Operation	
	Raw Naphtha 15-149°C	Depent. Naphtha
SR	15.7	11.8
HT	5.4	4.9
HC	3.2	2.9
		19.6

Average Composition of Combined Naphtha (Depent.)

		Vol %	P	O	N	A	°API	D86
SR	11.8	60.2	54.5	36.2	9.3		0.712	
HT	4.9	25.0	27.2	67.7	5.2		0.752	50% BP = 100°C
HC	2.9	14.8	29.7	66.7	3.6		0.746	
	19.6	100.0	44.0	48.6	7.4		0.727	

Motor Gasoline from Syncrude Naphtha

The following calculation provides an estimate of the reformer unit severity to produce typical unleaded regular motor gasoline.

Assumptions:

1) Reformate yield is 80 vol%

2) Only Regular Unleaded Gasoline is produced.

Octane required is $\frac{R+M}{2} = 87$

Calculation:

		Regular Unleaded Mix	
	Vol%	F1	F2
C_4	5	93.8	89.6
C_4-C_5 (Avg.)	25	78	77
Reformate	70	95	86
	100	90.7	84.0

$\frac{R+M}{2} = 87.4$

This calculation shows that 95 F1 octane reformate would be slightly more than sufficient to meet the blend requirement.

The octane values for the C_4-C_5 mix component were calculated as shown on Page 53.

The D86 distillation for the SR naphtha is shown in Figure 4.4.1 and the calculated D-186 for the depentanized naphtha is shown in Figure 4.4.2.

FIG 4.4.1 STRAIGHT RUN NAPHTHA

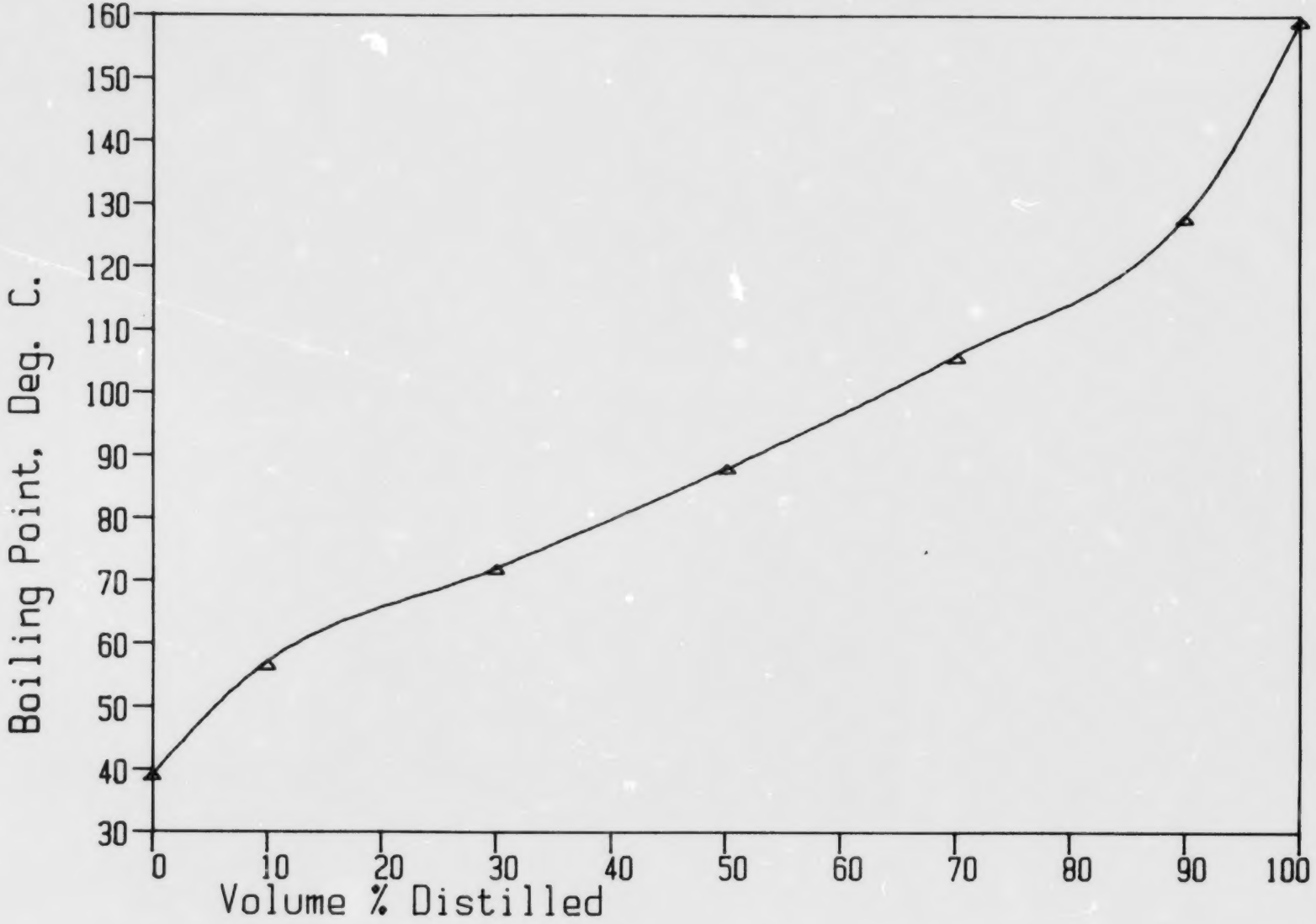
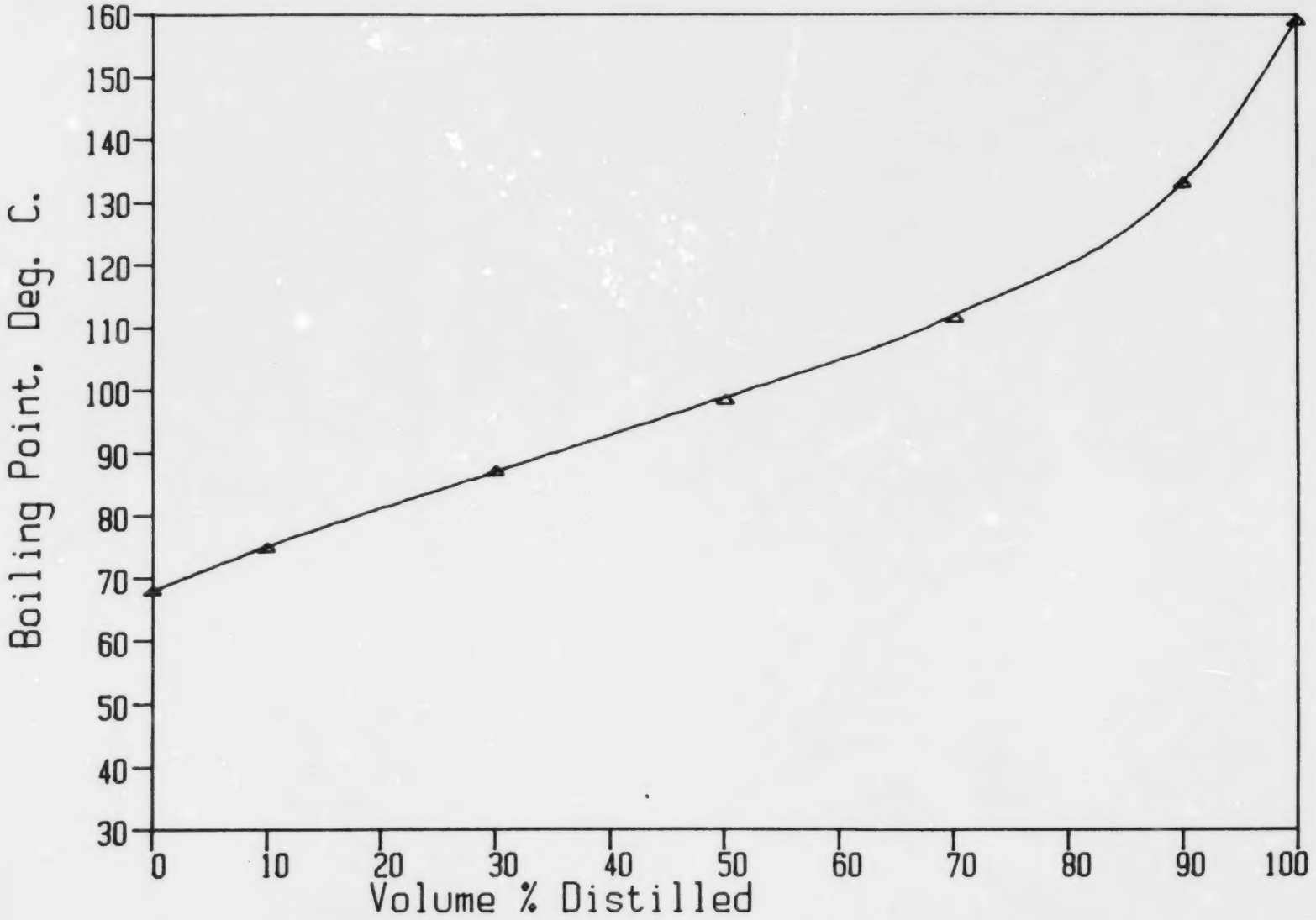


FIG. 4.4.2 DEPENTANIZED S. R. NAPHTHA (CALC.)



API DATA BOOK - Calculated Values

C_4 & C_5s	F1	F2
iC_4	100+	97.6
nC_4	93.8	89.6
iC_5	92.3	90.3
nC_5	61.7	62.6
$CycloC_5$	85.5	80.2

	HC Naphtha Fractions		
C_4 & C_5 Mix	Vol%	F1	F2
iC_4	5.0	5.0	4.9
nC_4	15.1	14.2	13.5
iC_5	35.0	32.3	31.6
nC_5	37.7	23.3	23.6
$CycloC_5$	7.2	6.2	5.8
	100.0	81.0	79.4

	SR Fractionated Naphtha		
C_4 & C_5 Mix	Vol%	F1	F2
iC_4	0.6	0.6	0.6
nC_4	12.0	11.3	10.8
iC_5	30.6	28.2	27.6
nC_5	47.9	29.6	30.0
$CycloC_5$	8.9	7.6	7.1
	100.0	77.3	76.1

	HT Naphtha Fractions		
C_4 & C_5 Mix	Vol%	F1	F2
iC_4	7	7.0	6.8
nC_4	30	28.1	26.9
iC_5	24	21.7	21.7
nC_5	35	21.6	21.9
$CycloC_5$	4	3.4	3.2
	100	81.8	80.5

5. CONCLUSIONS

At the end of the work in this project the following conclusions were drawn.

5.1 The project has shown that typical syncrude can be processed into large quantities of high quality diesel fuel. Good quality jet fuels were also demonstrated. The raw and produced naphthas in this work are acceptable as platforming feeds and may have use as solvent feedstocks as well.

5.2 The hydrotreatment of Syncrude Canada Ltd. source middle distillate and gas oil to produce quality fuel products requires significant severity of operation. Middle distillate treatment at LHSV 1.0 and 17.2 MPag pressure required 160 m^3 hydrogen/m^3 of feed to produce 45 cetane product. The conversion of gas oil into middle distillate by hydrocracking is a rigorous and expensive process in particular. The hydrocracking of gas oil at LHSV 0.5 and 20.3 MPag pressure required 225 m^3 hydrogen/m^3 of feed and produced 70 vol% yield middle distillate (on feed) with a 45 cetane number.

5.3 The commercial catalysts utilized in the study revealed a variety of performance levels in the treatment operations. Generally, the hydrotreatment performance, i.e. aromatic saturation varied with metals content and surface areas of the catalysts. The hydrocracking of gas oil was affected by the catalyst support material; the fluorided silica-alumina support promoted more cracking than the strictly alumina support investigated. The silica-alumina support was not desirable in middle distillate treatment due to yield loss. A high metals and high surface area alumina supported catalyst was selected for optimization in middle distillate treating. A catalyst with a high metals content and a high surface area silica-alumina support was chosen for gas oil hydrocracking.

5.4 The operating conditions which favoured aromatic saturation in middle distillate treating (on the selected catalyst) were high pressure and a moderate temperature (or severity) near 385°C for the space velocity chosen.

5.5 The operating conditions which favoured gas oil hydrocracking were found to be high pressure and a moderate temperature near 405°C (for the chosen catalyst). The temperature optimum was observed at the highest pressure investigated, 20.3 MPag. The gas oil conversion to lighter distillate continued to increase with increasing temperature but a net loss of middle distillate yield began to occur beyond the optimum temperature.

5.6 The observed yields from the processing on the original m^3 of syncrude were:

a) Stand-alone Middle Distillate Treatment

604.2 litres 149-385°C + 95.9 m^3 H_2	----- IBP-15°C	2.6 litres
	----- 15-149°C	31.1 litres
	----- 149-385°C	613.8 litres*
	TOTAL	647.5 litres

b) Stand-alone Gas Oil Hydrocracking

215.5 litres 385°C+ + 48.5 m^3 H_2	----- IBP-15°C	20.1 litres
	----- 15-149°C	27.7 litres
	----- 149-385°C	149.3 litres*
	----- 385°C+	58.7 litres
	TOTAL	255.7 litres

c) Blended Processing

1 m^3 Feed		Combined Production
23.3 litres	----- IBP-15°C	46.0 litres
157.0 litres	----- 15-149°C	215.8 litres
604.2 litres	----- 149-385°C	763.1 litres*
215.5 litres	----- 385°C+	58.7 litres
144.4 m^3 H_2	TOTAL	1083.6 litres

* Middle Distillate total includes Jet B blend off of 15°C-149°C naphtha according to the assumed sales ratio 25/4/1; Diesel/Jet A/Jet B.

5.7 The volume yield increase in the combined processing shows that the syncrude resource can be expanded using a relatively lower cost, resource (H_2 from N.G.) and by virtue of pricing differentials can create a net increase in liquid, value-added products at the same time. The processing offsets, however, are operating and capital costs of these higher severity processes.

5.8 With regard to product quality, we must conclude that the components can make acceptable fuel on a stand alone basis. Of course this conclusion is based on relatively fresh catalyst. As the catalyst ages due to coking or poisoning, we would expect a yield deterioration. In this study, the cost of the catalyst and catalyst life were not considered.

6. RECOMMENDATIONS

In light of the foregoing discussions and results, we recommend:

6.1 The study of catalyst life span, aging and potential poisons as a means of determining the overall costs/benefits associated with the severity of processing investigated herein. This includes the evaluation of yields and properties of effluent at both start of run and end of run.

6.2 The study of unconverted gas oil from the hydrocracking processing as a potential FCCU feedstock or as a feed for further hydrocracking. This should include the study of resultant product properties. An additional proposal could be the evaluation of two-stage vs. single-stage hydrocracking.

6.3 The optimization of diesel fuel production to suit the current CGSB specifications and the evaluation of the efficacy of cetane and combustion property improvers. This would potentially include the investigation of cetane response to operating severity in hydroprocessing.

6.4 The evaluation of the produced hydroprocessing naphthas in this study as potential sources for low aromatic solvent production.

6.5 The assessment of synthetic crudes produced from the next generation bitumen upgraders, in particular, the hydrogen addition types and the response of same source distillates to further hydroprocessing as performed in this work.

6.6 The assessment of synthetic crude distillates produced from delayed coking.

6.7 Combustion tests for the Jet fuels are recommended. These are highly naphthenic fuels, and the EMR Scientific Authority has suggested they might regress to aromatics in the combustion chamber.

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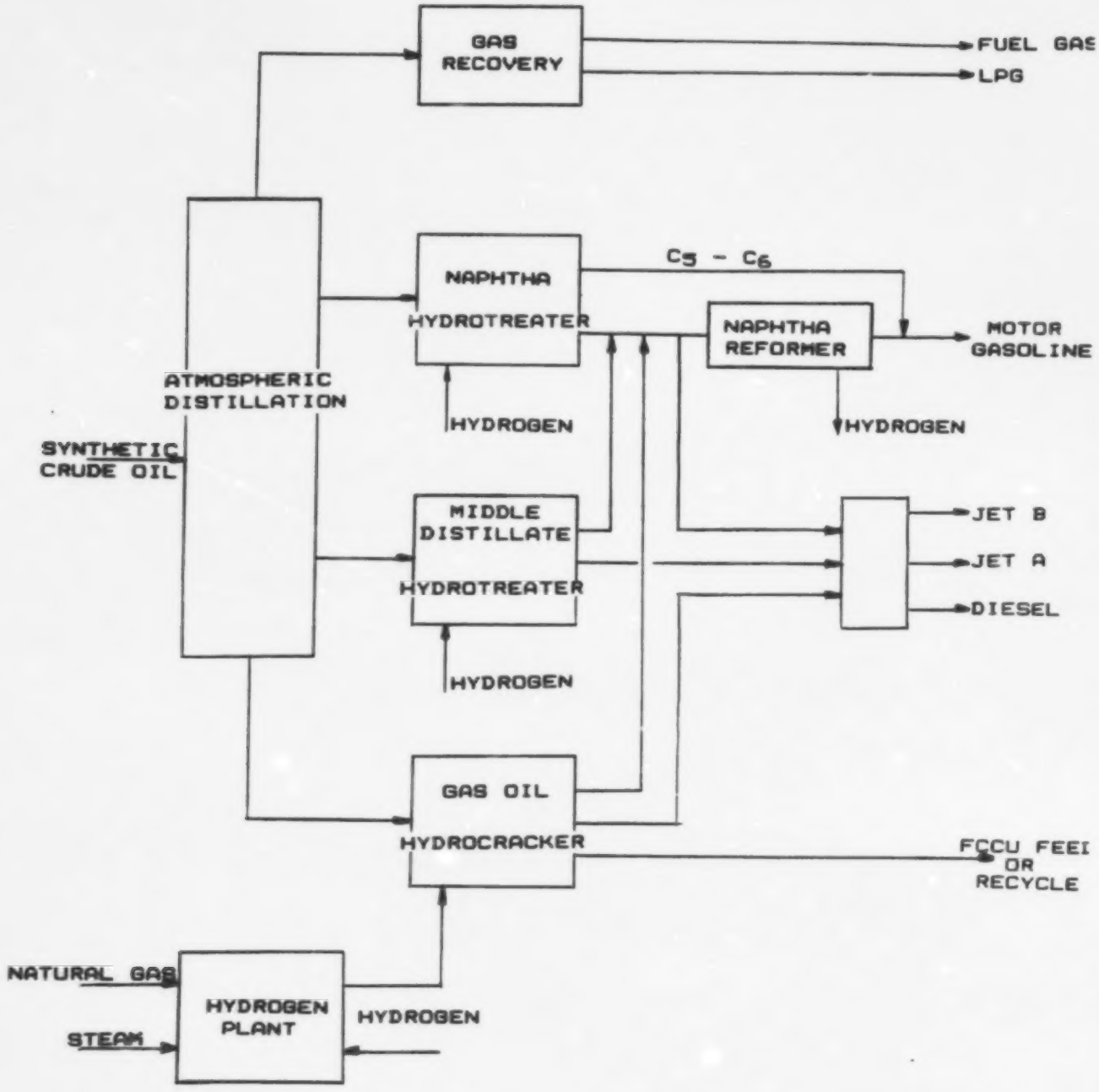
Figure 2.1.1 Process Configuration



Project Overview

FIGURE 3.1.1

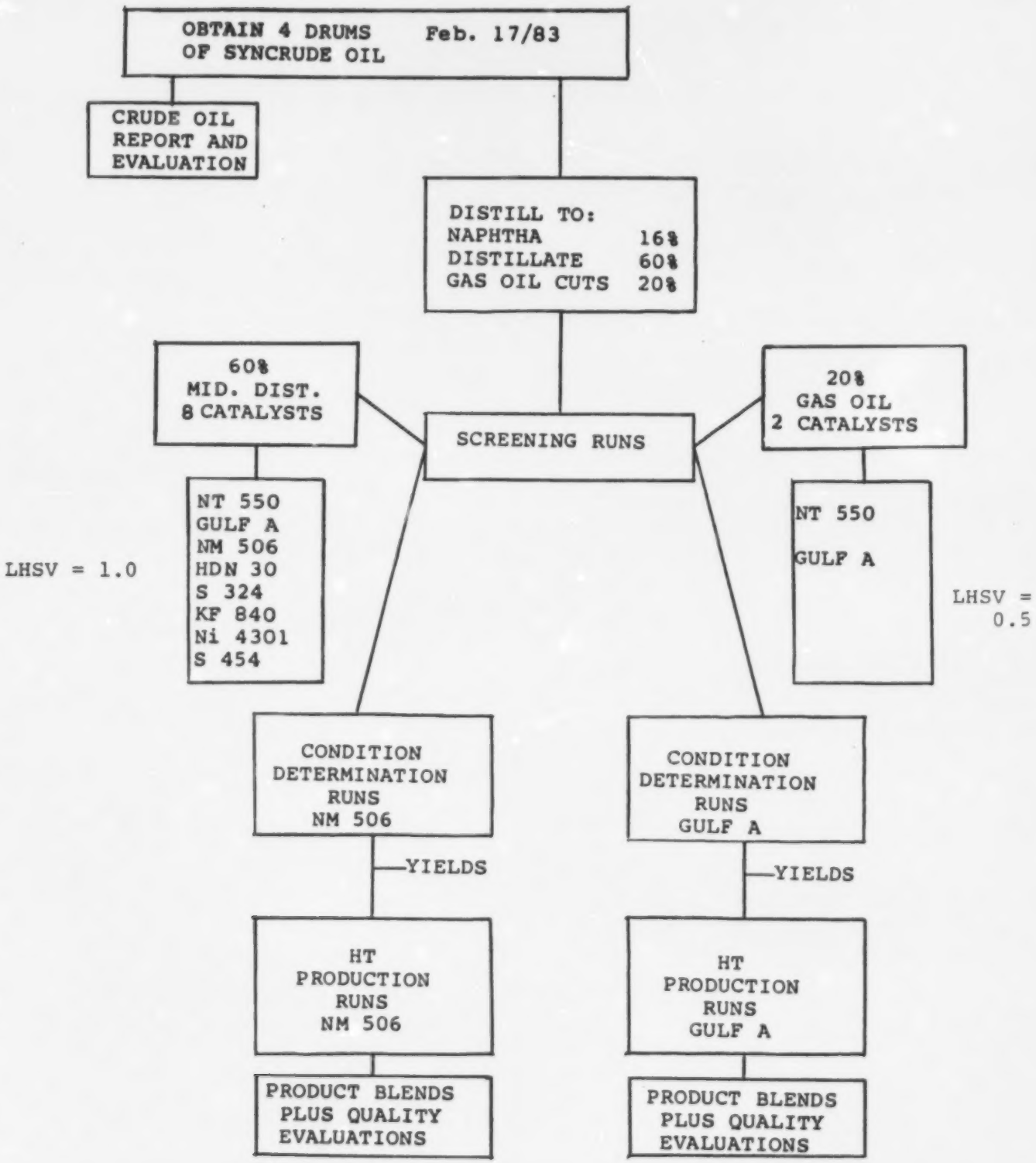


Figure 3.2.1 P-12 Hydrotreater Diagram

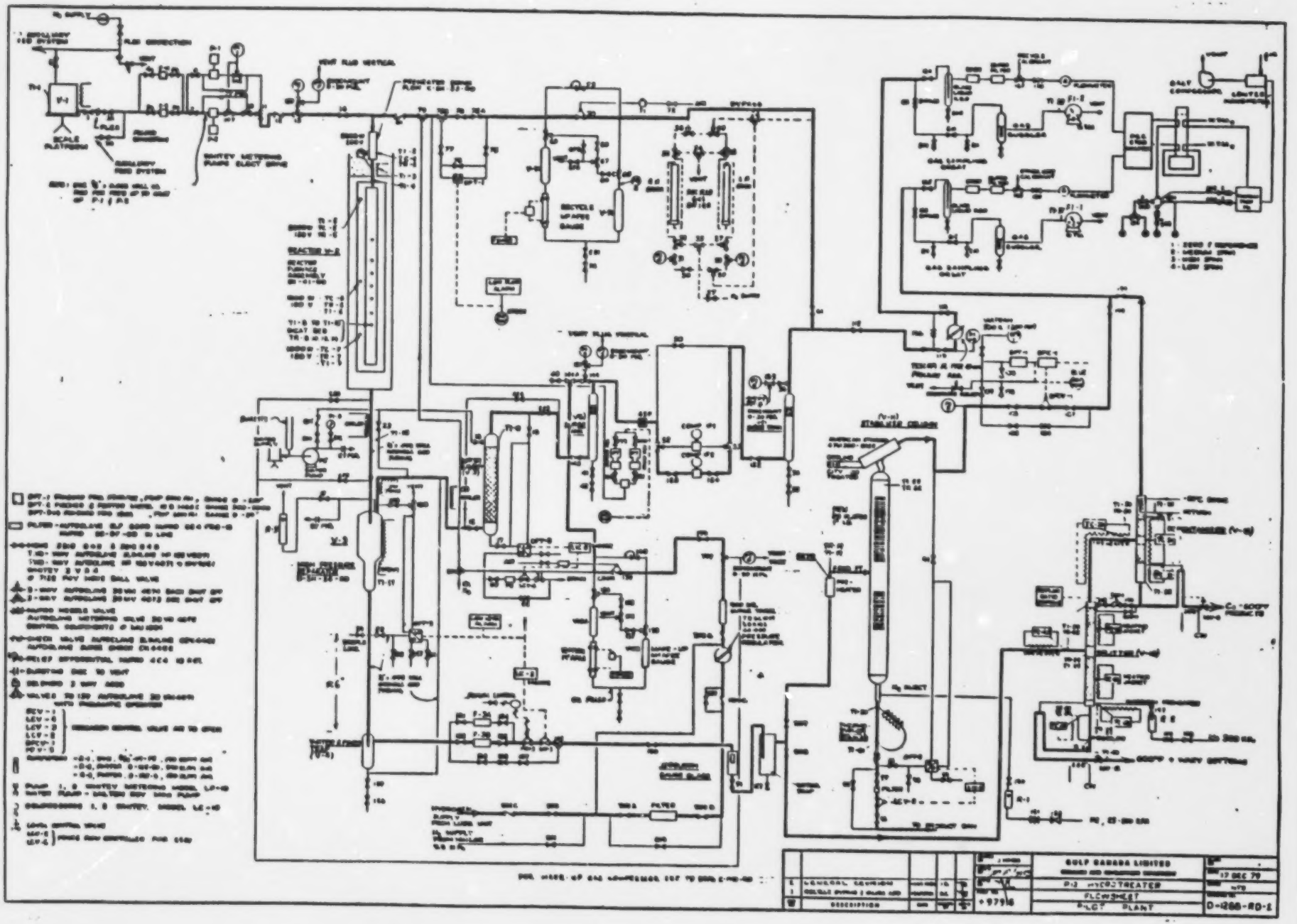


Figure 3.2.2 P-16 Hydrotreater Diagram

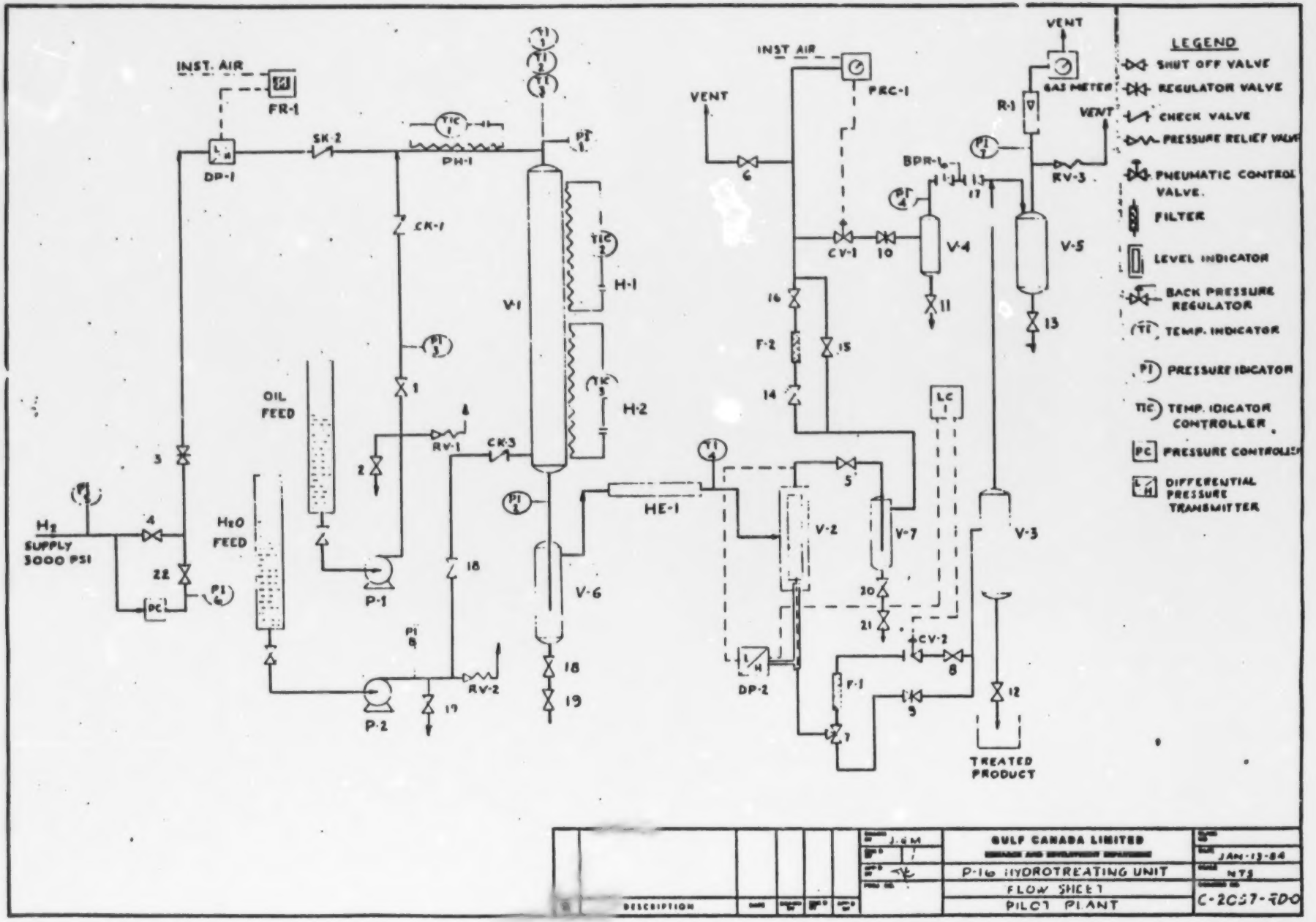


Figure 4.1.1 Crude Assay True Boiling Point Curve

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RESEARCH AND DEVELOPMENT DEPARTMENT PAGE 1

CRUDE SYNCRUDE SAMPLE NO. D83-12-5 PROJECT NO. A64-83-65

GRAVITY °API 31.6

SPECIFIC GRAVITY 60/60 _____

DENSITY kg/L @ 15°C 0.8670

TRUE BOILING POINT CURVE

DENSITY MID % CURVE

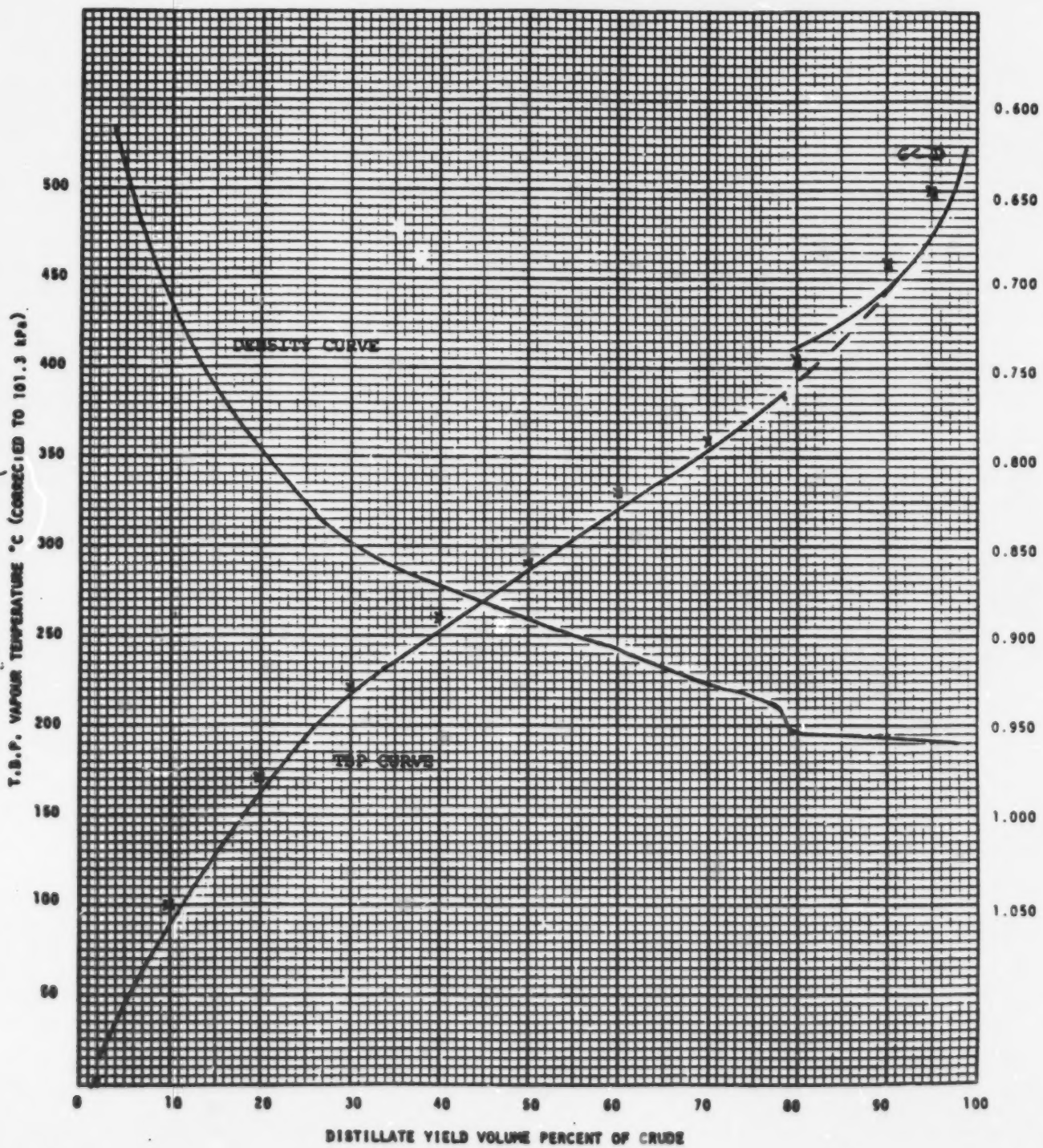

FIG. 4.2.1A AROMATIC CONVERSION Ni/W CAT. vs WABT

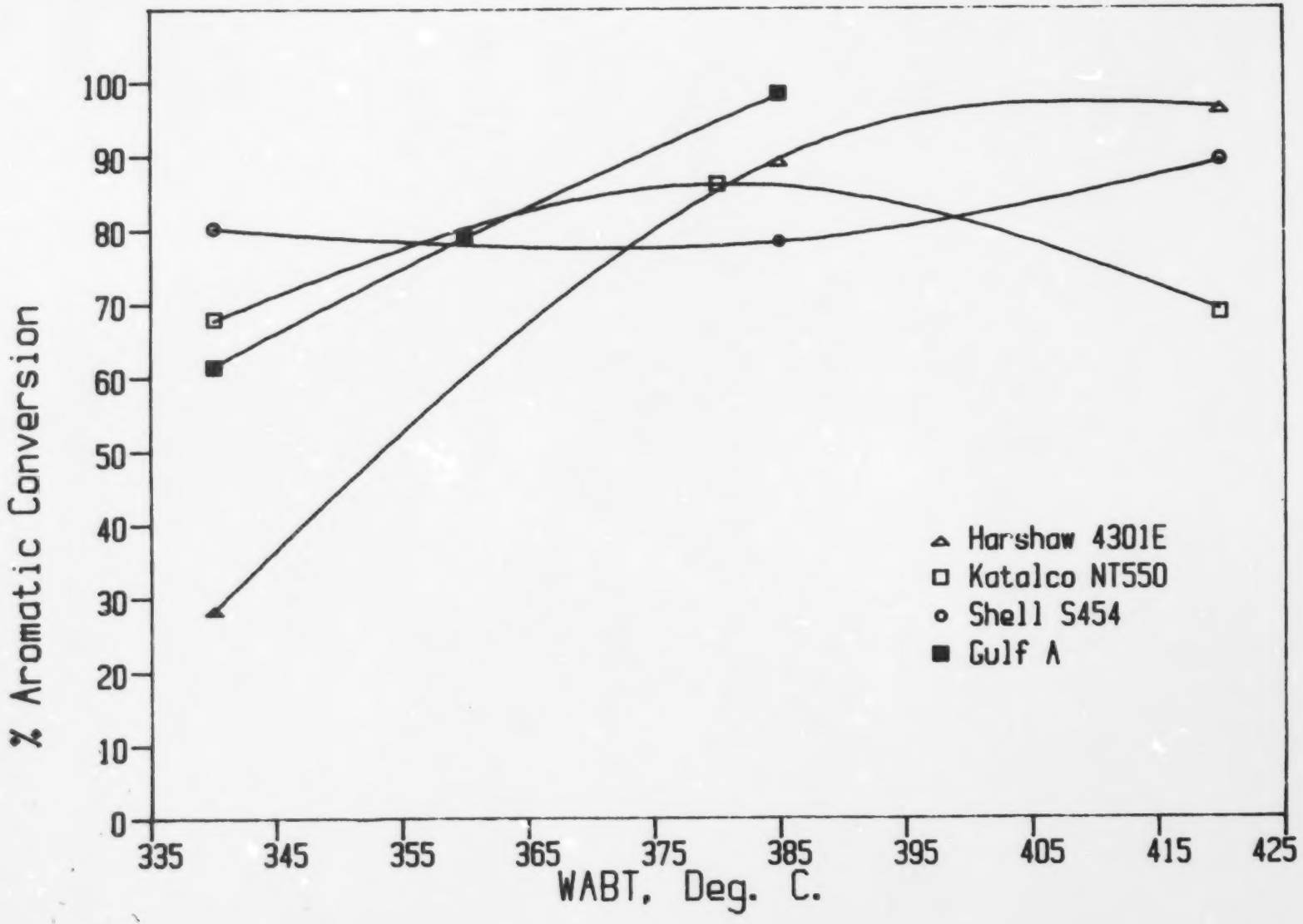

FIG. 4.2.1B AROMATIC CONVERSION Ni/Mo CAT. vs WABT

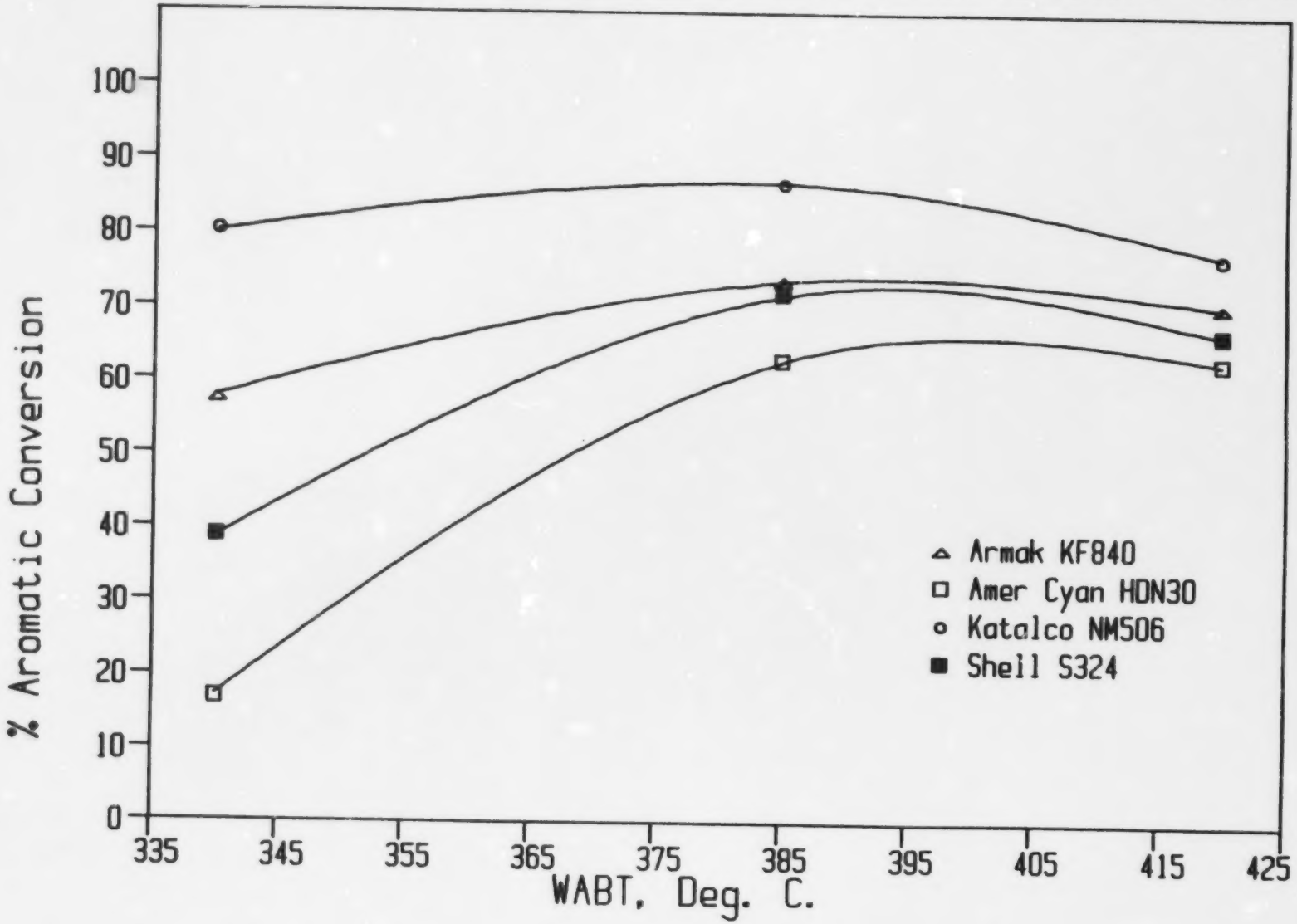

FIG. 4.2.2A PRODUCT MOLECULAR WEIGHT vs WABT

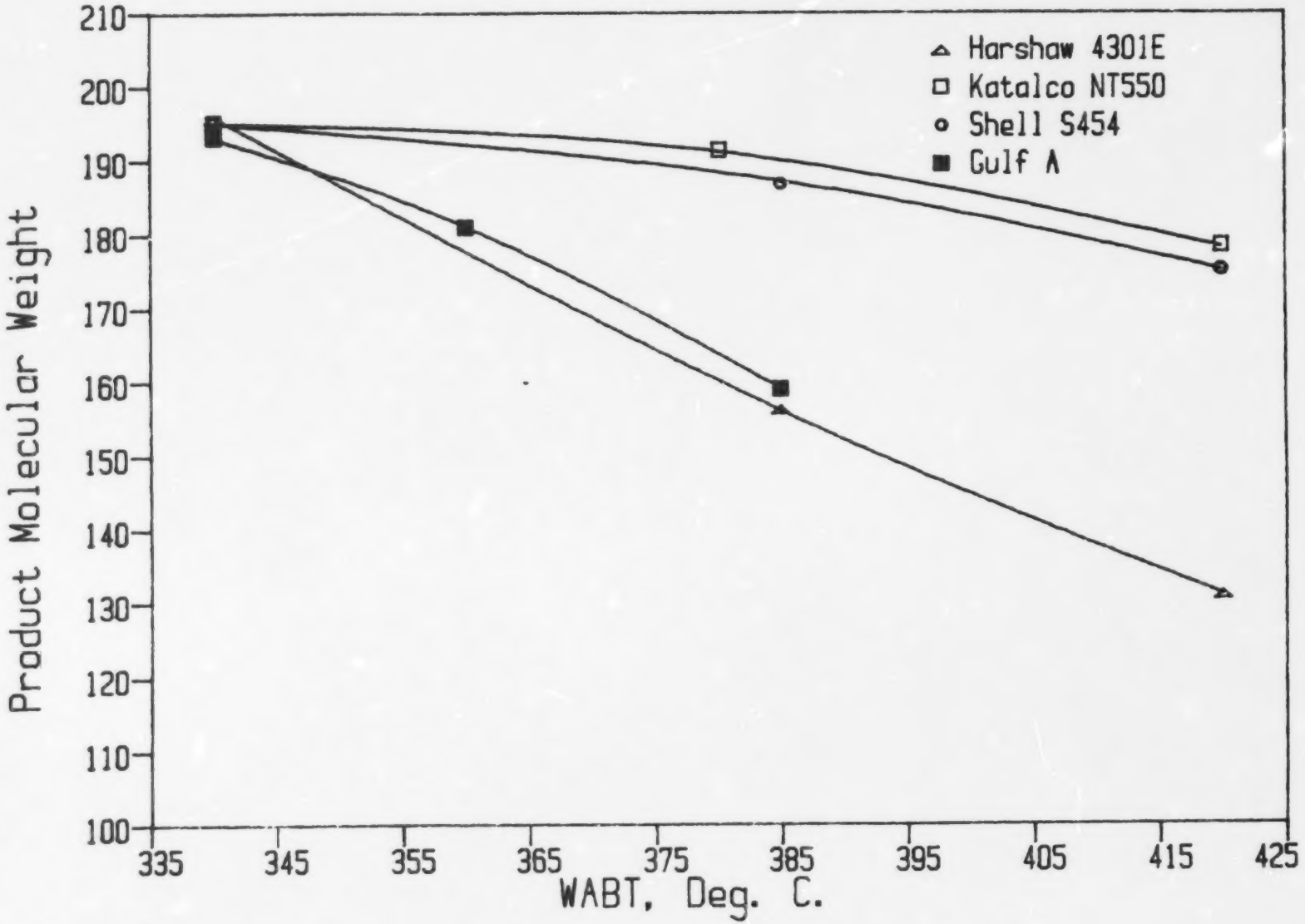

FIG. 4.2.2B PRODUCT MOLECULAR WEIGHT vs WABT

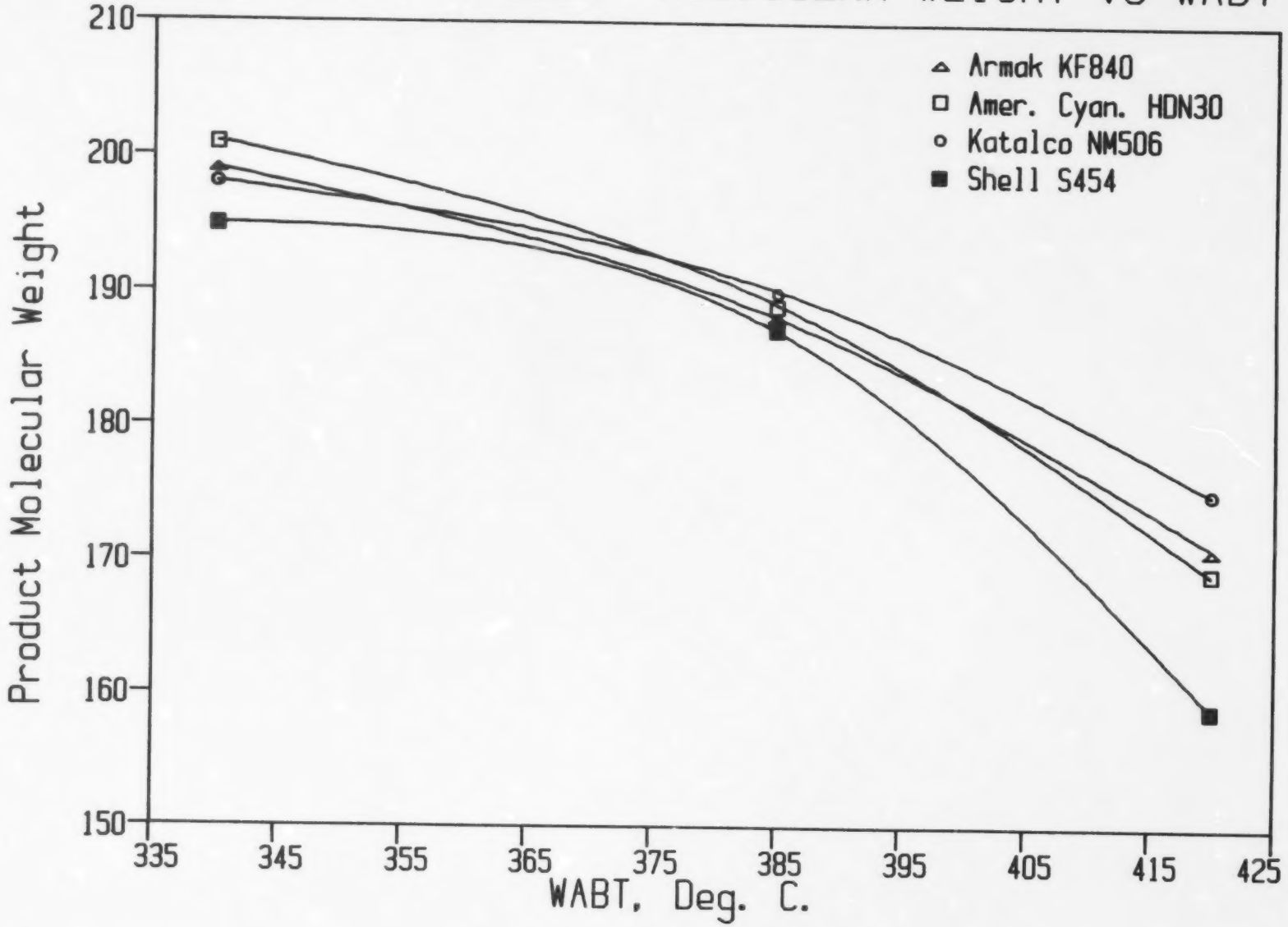

FIG. 4.2.3 WT. % Ni vs OPT. AROMATIC CONV.

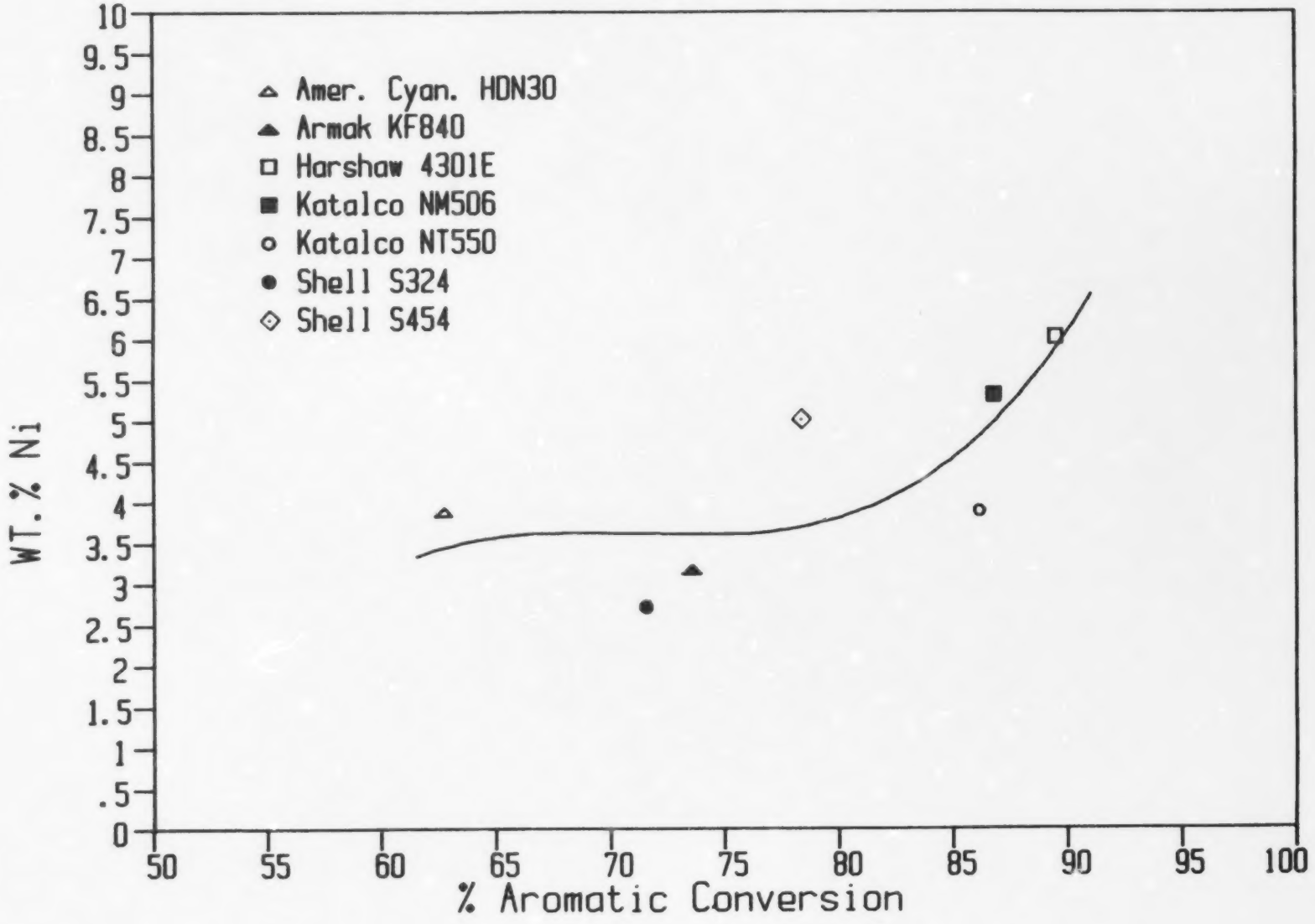

FIG. 4.2.4 SURFACE AREA vs OPT. AROMATIC CONV.

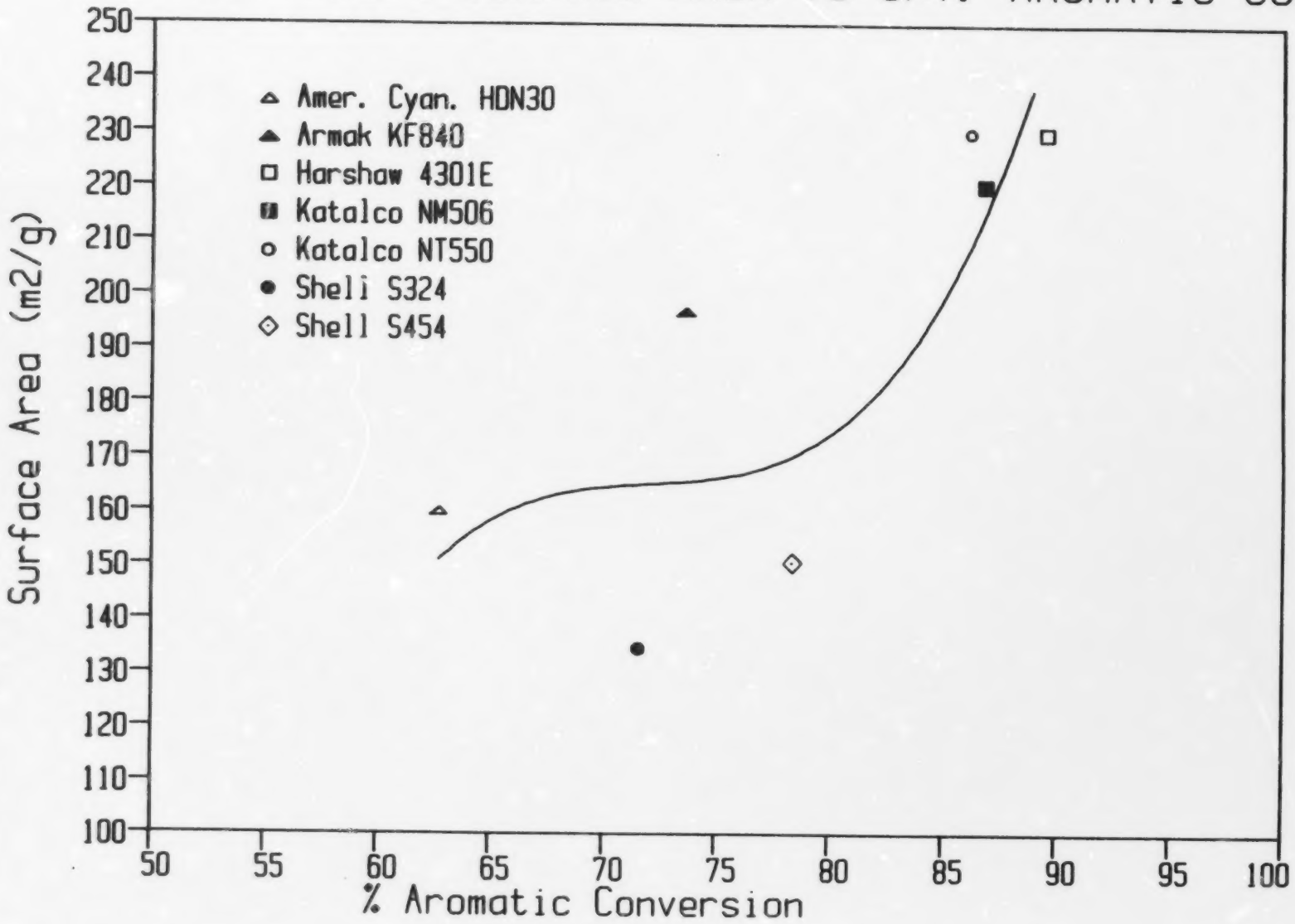

FIG. 4.2.5 NM506 AROMATIC CONVERSION vs WABT

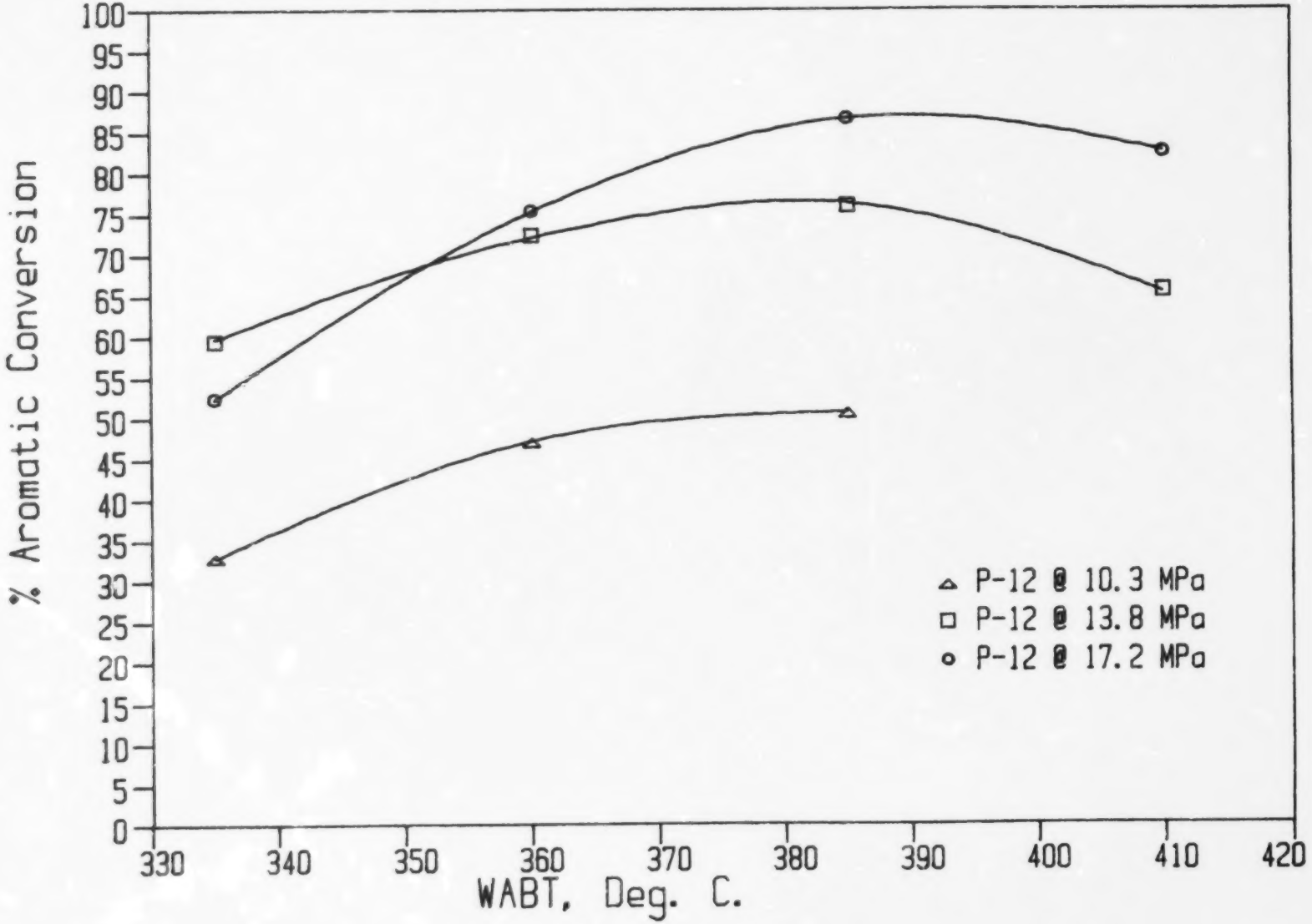

FIG. 4.2.6 AROMATIC CONVERSION vs PRESSURE

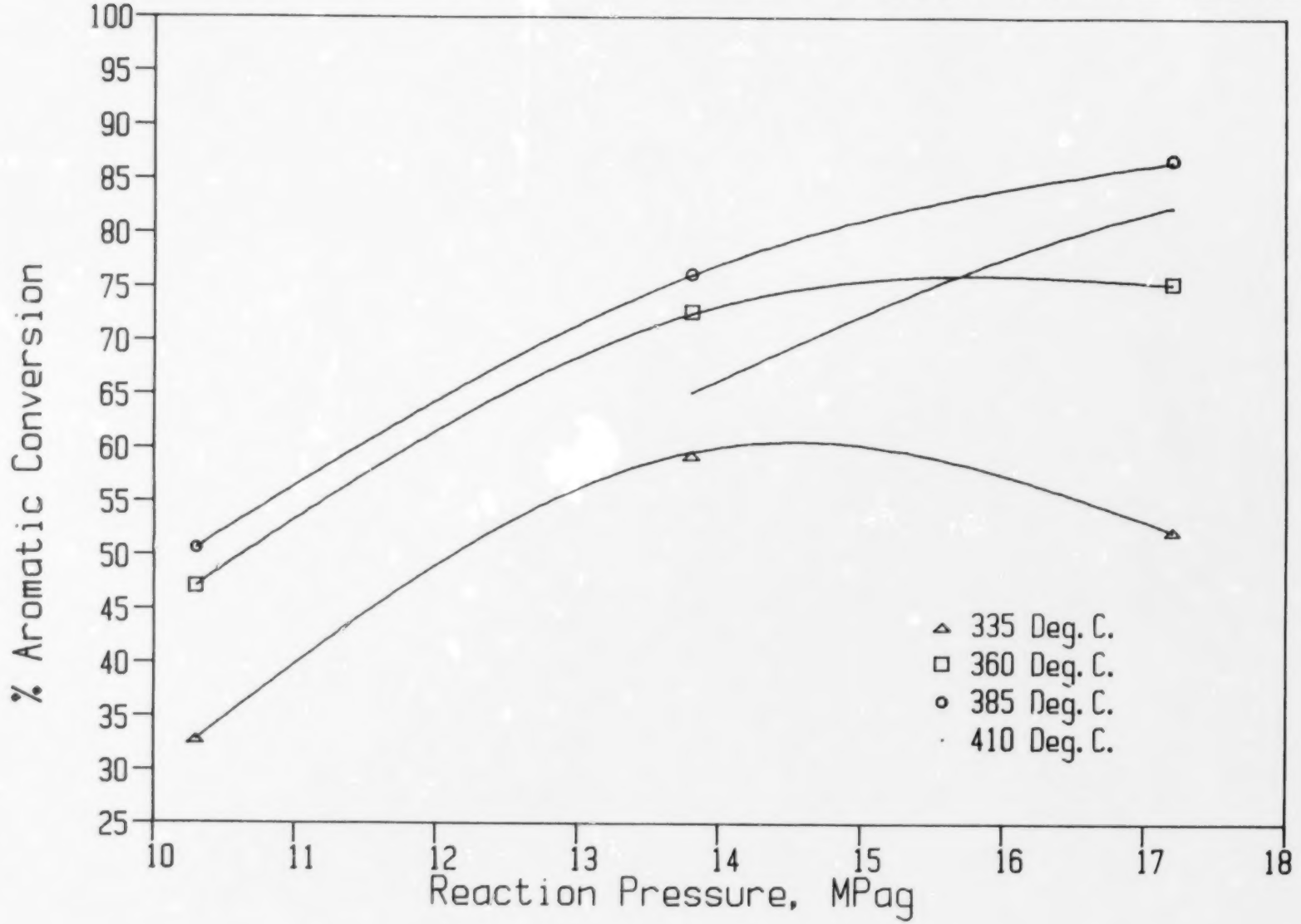

FIG. 4.3.1A GAS OIL CONVERSION vs WABT

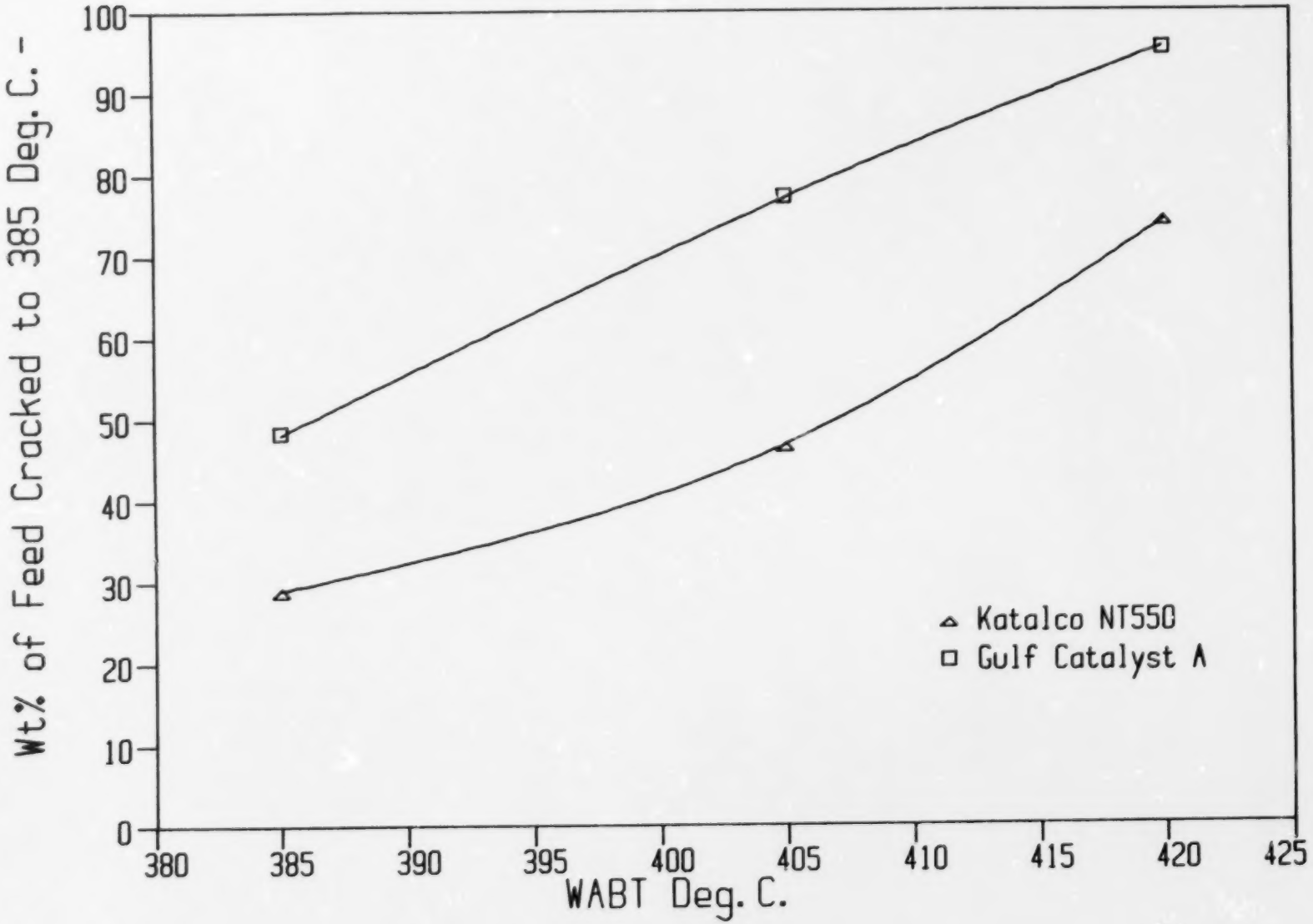

FIG. 4.3.1B GAS OIL CONVERSION vs WABT

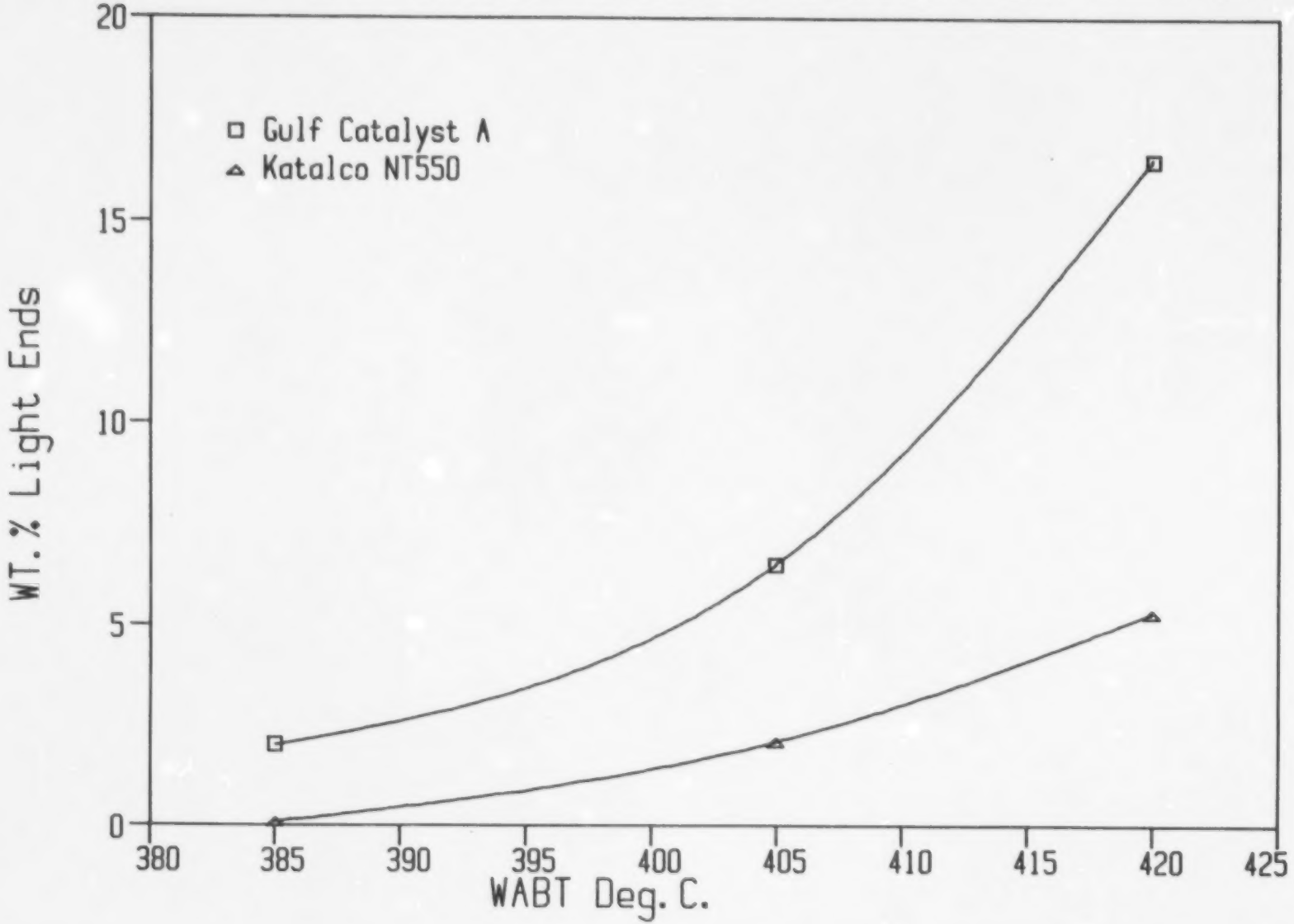

FIG. 4.3.2 MIDDLE DISTILLATE YIELD vs REACTOR TEMP

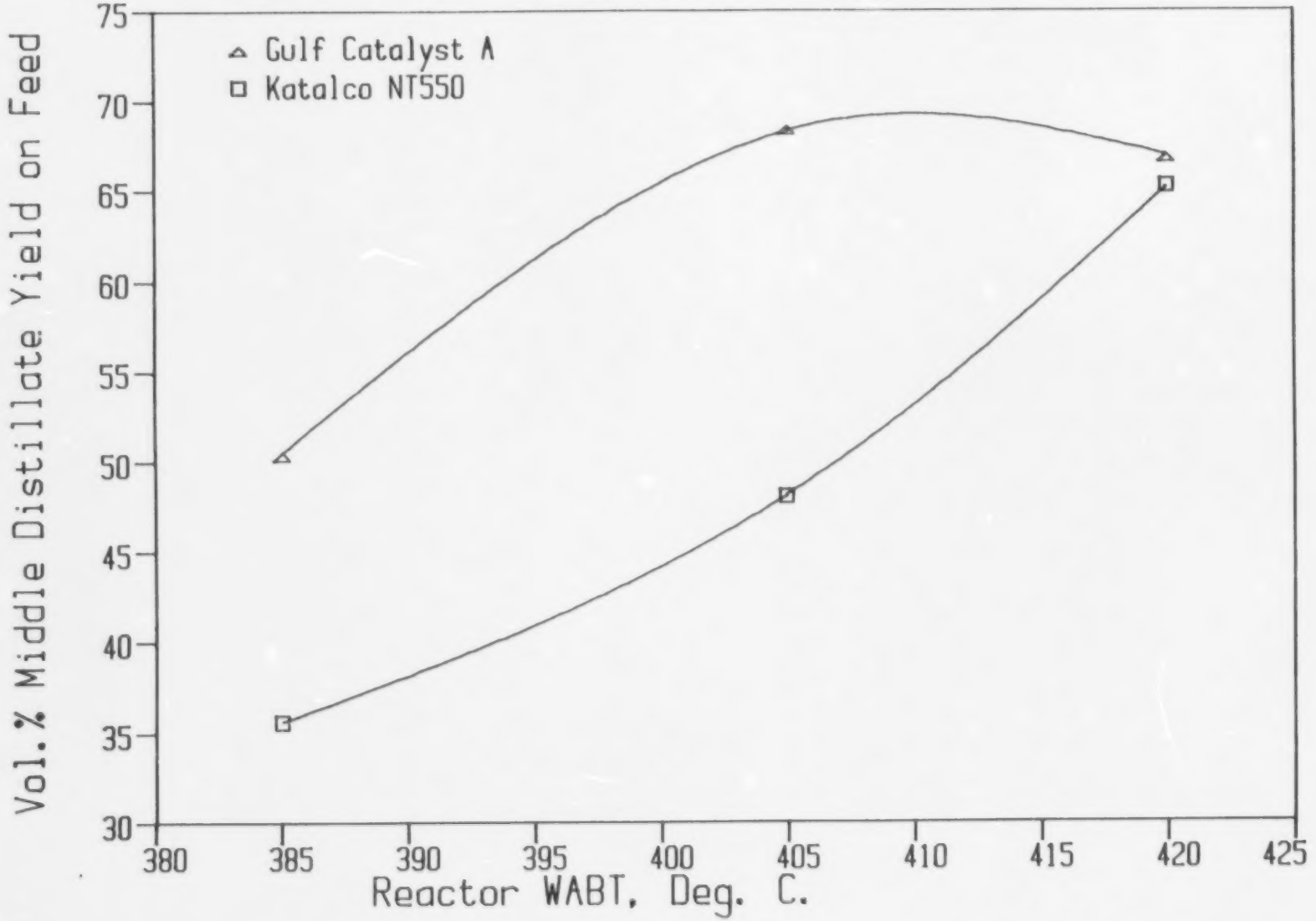

FIG. 4.3.3 MIDDLE DISTILLATE YIELD vs REACTOR TEMP

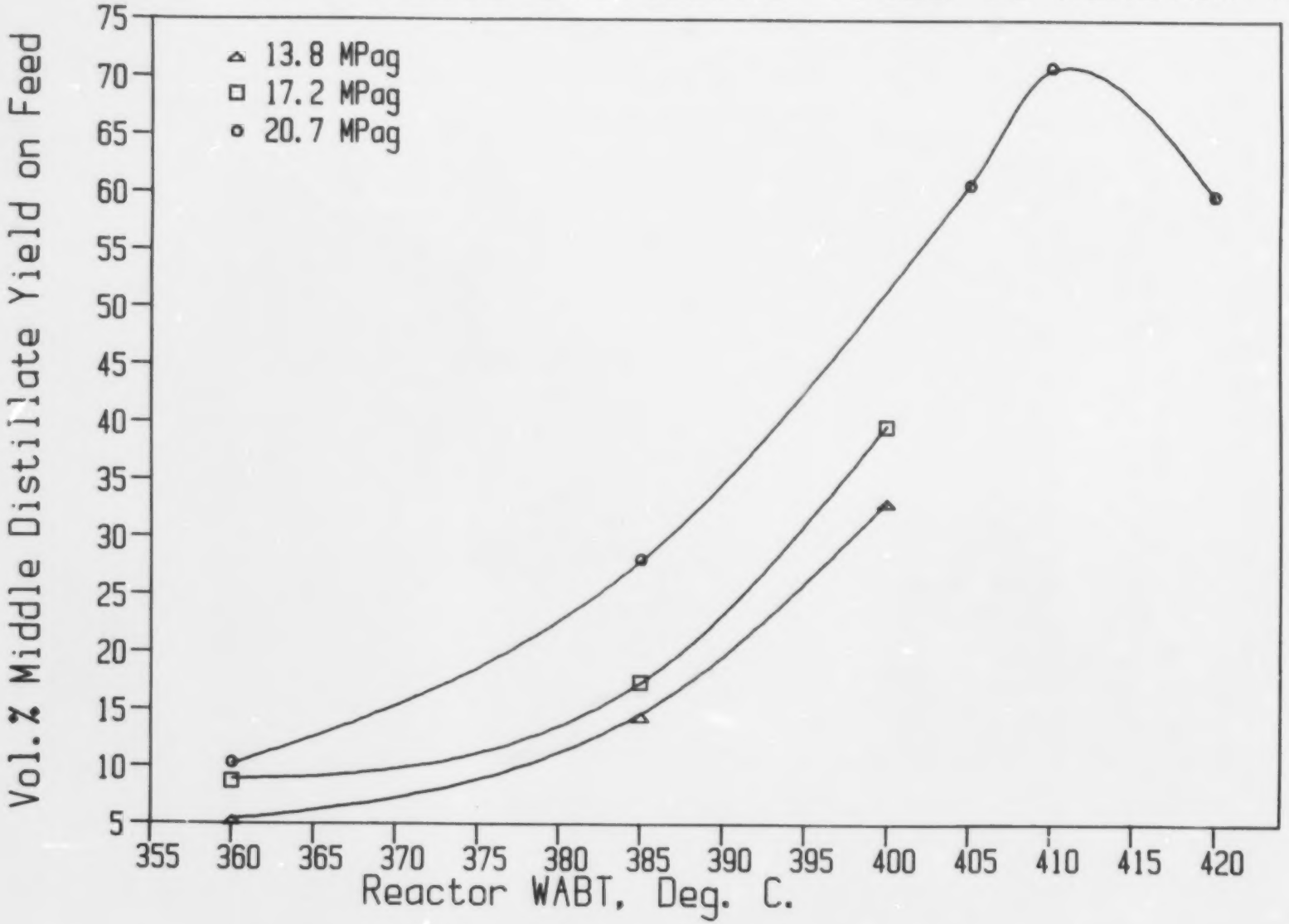

TABLE 3.4.1 SUMMARY OF CGSB JET A SPECIFICATIONS

		Jet A Min.	Jet A Max.	Jet A-1 Min.	Jet A-1 Max.	Jet A-2 Min.	Jet A-2 Max.	Test Method / Méthode d'essai
5.1	Flash Point, °C	38	--	38	--	33	--	ASTM D 56 or/ou D 3243
5.2	Freezing Point, °C							
	(Note 1) Summer (May 1 – Oct. 31)	--	-40 (Note/ Remarque 2)	--	-47	--	-40	ASTM D 2386
	Winter (Nov. 1 – Apr. 30)	N/A		--	-47	--	-47	
5.3	Kinematic Viscosity at -20°C (mm^2/s) (Note 3)			--	8			ASTM D 445
5.4	Distillation							ASTM D 86
5.4.1	10% Recovered Point, °C			--	204			
5.4.2	50% Recovered Point, °C			Report/Indiquer				
5.4.3	90% Recovered Point, °C			Report/Indiquer				
5.4.4	Final Boiling Point, °C			--	300			
5.4.5	Distillation Residue, % vol			--	1.5			
5.4.6	Distillation Loss, % vol			--	1.5			
5.5	Density, kg/m^3 at 15°C	775		775	840			ASTM D 1298
5.6	Sulfur, % mass			--	0.2			ASTM D 1266 or/ou D 2622 (Note/ Remarque 4)
5.7	Sulfur, Mercaptan, % mass (Note 5)			--	0.003			ASTM D 3227
5.8	Corrosion (Copper Strip-Bomb) 2 h at 100°C			--	No./Nº 1			ASTM D 130
5.9	Corrosion (Silver Strip) (Note 6)			--	No./Nº 1			IP 227
5.10	Copper Content, mg/L (Note 7)			--	0.15			3-GP-0/131.1
5.11	Water Separation Index (Modified)			75	--			ASTM D 2550 or/ou D 3602 or/ou D 3948
5.12	Water Reaction							ASTM D 1094
	Separation rating			--	2			
	Interface rating			--	1 b			

		Jet A, Jet A-1 and/et Jet A-2 Min.	Jet A, Jet A-1 and/et Jet A-2 Max.	Test Method / Méthode d'essai
5.13	Net Heat of Combustion (Lower Heat Value), MJ/kg	42.8	--	ASTM D 1405 or/ou D 3338 (Note/ Remarque 8)
5.14	Aromatics, % vol (Note 9)	--	22.0	ASTM D 1319
5.15	Combustion Properties One of the following:			
	a. Luminometer number (Note 10)	45	--	ASTM D 1740
	b. Smoke Point, mm	25	--	ASTM D 1322
	c. Smoke Point, mm, plus Naphthalenes, % vol (Note 11)	20 --	-- 3	ASTM D 1322 ASTM D 1840
5.16	Total Acidity, mg KOH/g	--	0.1	ASTM D 974
5.17	Particulate Matter, mg/L, at time of delivery to:			ASTM D 2276 (Note/ Remarque 12)
	a. Aircraft and refuellers	--	0.44	
	b. Purchasers storage	--	2.2	
5.18	Electrical Conductivity, pS/m at point, time and temperature of delivery to purchaser (par. 5.20.5)	50	450	ASTM D 2624
5.19	Thermal Stability – The thermal stability shall be measured using either of the methods described in par. 5.19.1 or 5.19.2. It is anticipated that the Alcor JFTOT will eventually replace the ASTM-CRC Fuel Coker. Meanwhile both tests should be reported whenever possible. *(At present ASTM D 1660 will be the referee test method in the event of dispute).*			
5.19.1	*ASTM-CRC Fuel Coker* – The change in pressure drop across the filter shall not exceed 10 kPa and the preheater deposit shall rate less than 3 when tested in accordance with ASTM D 1660, subject to the following conditions:			
	Preheater temperature	149 ± 3°C		
	Filter temperature	204 ± 6°C		
	Fuel flow	2.7 ± 0.05 kg/h		
	Test time	300 min		
5.19.2	*Alcor Jet Fuel Thermal Oxidation Tester (JFTOT)* – The pressure drop across the filter shall not exceed 3.4 kPa and the heater tube deposit visual rating shall be less than 3 (blue or grey deposits shall be cause for rejection) when tested in accordance with ASTM D 3241 and subject to the following conditions:			
	Maximum heater tube temperature	260°C		
	Fuel system pressure	3.45 MPa		
	Fuel flow rate	3 mL/min		
	Test time	150 min		

Testing facilities equipped with Mark 8A tube deposit rater should record and report spun ratings as well

TABLE 3.4.2 SUMMARY OF CGSB JET B SPECIFICATIONS

		JET B Min.	JET B Max.	F-40 Min.	F-40 Max.	Test Method Méthode d'essai
5.1	Freezing Point, °C NOTE: The –58°C maximum freezing point of the F-40 grade fuel shall be in effect until December 31, 1980 at which time it shall be brought in line with the JET B freezing point of –51°C.	–	–51	–	–58	ASTM D 2386
5.2	Reid Vapor Pressure, kPa	–	21	14	21	ASTM D 323
5.3	Electrical Conductivity, pS/m at point, time and temperature of delivery to purchaser (par. 5.21.5)	50	450	100	600	ASTM D 2624
5.4	Existent Gum (Air or Steam Jet) mg/100 mL	–	–	–	7.0	ASTM D 381
5.5	Total Acidity, mg KOH/g		–		.1	ASTM D 974
5.6	Distillation					ASTM D 86
5.6.1	Initial Boiling Point, °C		Report/*Indiquer*			
5.6.2	10% Recovered Point, °C		Report/*Indiquer*			
5.6.3	20% Recovered Point, °C				143	
5.6.4	50% Recovered Point, °C				188	
5.6.5	90% Recovered Point, °C				243	
5.6.6	End Point, °C		Report/*Indiquer*			
5.6.7	% Recovered at 204°C		Report/*Indiquer*			
5.6.8	Distillation Residue, % vol		–		1.5	
5.6.9	Distillation Loss, % vol		–		1.5	
5.7	Density, kg/L at 15°C		0.750		0.801	D 1298
5.8	Aromatics, % vol		–		25.0	D 1319
5.9	Olefins, % vol		–		5.0	D 1319
5.10	Sulfur, % mass		–		0.4	ASTM D 1266* or/*ou* D 2622
5.11	Sulfur, Mercaptan, % mass (Note 1)		–		.003	ASTM D 1219 or/*ou* D 3227**
5.12	Corrosion (Copper Strip Bomb) 2 h at 100°C		–		No./*N°* 1	ASTM D 130
5.13	Corrosion (Silver Strip) (Note 2)		–		No./*N°* 1	IP 227
5.14	Copper Content, mg/L (Note 3)		–		0.15	3-GP-0/131.1
5.15	Water Separation Index (Modified)	75		–		ASTM D 2550 or/*ou* D 3602
5.16	Water Reaction					ASTM D 1094
	Separation rating		–		2	
	Interface rating		–		1 b	
5.17	Net Heat of Combustion (Lower Heat Value) MJ/kg	42.8		–		ASTM D 1405 or/*ou* D 2382***
5.18	Combustion Properties: one of the following					
	1. Luminometer number,	45		–		ASTM D 1740
	or 2. Smoke Point, mm,	25		–		ASTM D 1322
	or 3. Smoke Point, mm, plus Naphthalenes, % vol	20 –		– 3		ASTM D 1322 ASTM D 1840
	or 4. Smoke Volatility Index	52		–		(Note 4/ *Remarque 4)*
5.19	Particulate Matter, mg/L at time of delivery to:					ASTM D 2276 (Note 5/ *Remarque 5)*
	a. Aircraft and refuellers		–		0.44	
	b. Purchasers storage		–		2.2	

5.20 **Thermal Stability** – The thermal stability shall be measured using either of the methods described in par. 5.20.1 or 5.20.2. It is anticipated that the Alcor JFTOT will eventually replace the ASTM-CRC Fuel Coker. Meanwhile both tests should be reported whenever possible. *(At present ASTM D 1660 will be the referee test method in the event of dispute).*

5.20.1 *ASTM-CRC Fuel Coker* – The change in pressure drop across the filter shall not exceed 10 kPa and the preheater deposit shall rate less than 3 when tested in accordance with ASTM D 1660, subject to the following conditions:

Preheater temperature	= 149 ± 3°C
Filter temperature	= 204 ± 6°C
Fuel flow	= 2.7 ± 0.05 kg/h
Test time	= 300 min

5.20.2 *Alcor Jet Fuel Thermal Oxidation Tester (JFTOT)* The pressure drop across the filter shall not exceed 3.4 kPa and the heater tube deposit visual rating shall be less than 3 (blue or grey deposits shall be cause for rejection) when tested in accordance with ASTM D 3241 and subject to the following conditions:

Maximum heater tube temperature	= 260°C
Fuel system pressure	= 3.45 MPa
Fuel flow rate	= 3 mL/min
Test time	= 150 min

Testing facilities equipped with Mark 8A tube deposit rater should record and report spun ratings as well as visual ratings.

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*ASTM D 1266 shall be used as the referee method in the event of a dispute.

**ASTM D 3227 shall be used as the referee method in the event of a dispute.

***ASTM D 2382 shall be used as the referee method in the event of dispute.

Note 1. The mercaptan sulfur determination may be waived if the fuel is considered "sweet" by the Doctor test described in ASTM D 484, section 6.2 for Hydrocarbon Dist llate Solvents.

Note 2. Purchaser Option (par. 7.1)

Note 3. The Copper Content requirement applies only for fuels not subject to the "Copper Systems" requirement.

Note 4. Smoke Volatility Index (SVI) = Smoke point (mm) + 0.42 (Naphthalene % recovered at 204°C).

Note 5. A minimum of 4 L shall be filtered.

TABLE 3.4.3 SUMMARY OF CGSB DIESEL SPECIFICATIONS

		TYPE AA	TYPE A	TYPE B	Method Méthode ASTM
Flash Point Point d'éclair	°C, Min.	40	40	40	D 3828 D 93 (Note 4) (Remarque 4)
Cloud Point Point de trouble	°C, Max.	-48	-34	(Note 5) (Remarque 5)	D 2500
Pour Point Point d'écoulement	°C, Max.	-51	-39	(Note 5) (Remarque 5)	D 97 (Note 6) (Remarque 6)
Kinematic Viscosity at 40°C, mm2/s (cSt) (Note 7) Viscosité cinématique à 40°C, mm2/s (cSt) (Remarque 7)	Min. Max.	1.2 -	1.3 4.1	1.4 4.1	D 445
Distillation: 90% Recovered Récupéré à 90%	°C, Max.	290	315	360	D 86
Water and Sediment, % vol. Eau et sédiment, % en vol.	Max.	0.05	0.05	0.05	D 1796 (Note 8) (Remarque 8)
Total Acid Number Indice d'acide total	Max.	0.10	0.10	0.10	D 974
Sulphur, % mass (Note 9) Soufre, % en masse (Remarque 9)	Max.	0.2	0.5	0.7	D 1552 (Note 10) (Remarque 10)
Copper Corrosion 3 h at 100°C Corrosion du cuivre 3 h à 100°C	Max.	No./No 1	No./No 1	No./No 1	D 130
Carbon Residue (Ramsbottom), on 10% bottoms, % mass (Note 11) Résidue de carbone (Rambsbottom) sur 10% de résidus, % en masse (Remarque 11)	Max.	0.15	0.15	0.20	D 524
Ash, % mass Cendre, % en masse	Max.	0.01	0.01	0.01	D 482
Ignition Quality, Cetane No. (Note 12) Aptitude à l'allumage, indice de cétane (Remarque 12)	Min.	40	40	40	D 613

Table 4.1.1 Crude Assay Data Summary

GULF CANADA LIMITED

RESEARCH AND DEVELOPMENT DEPARTMENT
PRODUCT ANALYSIS

SAMPLE NO.: D83-12-5
STILL:
CRUDE: Syncrude

Fraction	METHOD REFERENCE	CRUDE	NAPHTHA	DISTILLATE	GAS OIL	RESID	385+ GAS OIL
Range (TBP) °C			15-149	149-385	385-524	524+	385+
Range from Vol. %			2.33	18.03	78.45	98.84	78.45
to Vol. %			18.03	78.45	98.84	100.00	100.00
Yield Vol. %			15.70	60.42	20.39	1.16	21.55
Gravity °API	D1298/GCM 243	31.6	67.4	28.7		14.8	16.3
Density @ 15°C		0.8678	0.7114	0.8833		0.9676	0.9549
Sulphur %	XRF	0.15		0.10		0.41	0.40
	GCM 100						
	GCR 54D	0.12	0.02				
Ramsbottom C.R. %	D524					2.16	0.26
Total Nitrogen ppm	(GCM 176) D3431	459	<1	167		(2150)	1510
Cloud Pt. °C	D2500			-33			Too Dark
Pour Pt. °C	D97			Below -51			6
Freeze Pt.	D2386						
Smoke Pt.	D1332						
Luminometer No.	D1740						
Cetane No. Clear	D613			30			
Flash Pt. °C	D93			72			
Distillation, °C	D86/D1160						
IBP	(GCM 379)	35	39	173	(361)		396
10%		109	57	206	(392)		408
30%		232	72	247	(416)		419
50%		296	88	281	(437)		437
70%		353	106	310	(461)		458
90%			128	341	(496)		501
FBP			159	357	(551)		570
Recovery, %			99.0	99.0			99.0
Viscosities	D445						
cSt @ 25°C		4.272					
cSt @ 40°C		3.096		6.590			190.34
cSt @ 100°C						63.53	11.15
VI							-15
Aniline Pt. °C	D611						
Nickel ppm	GCM 239				<0.05	0.82	0.15
Vanadium ppm	GCM 239				<0.05	0.51	0.07
Copper ppm	GCM 195				<0.05	2.16	0.19
Iron ppm	GCM 195				0.86	53	4.78
Lt. Ends Vol %	GCM 147/223						
C_2		—	—				
C_3		0.18	—				
iC_4		0.56	0.15				
nC_4		2.15	3.21				
iC_5		1.24	7.75				
nC_5		1.90	12.30				
C_6+		93.97	76.59				
				Dist.			GCM 33
HC Composition, Vol.%	GCR 162			LRMS Mass%			LRMS Mass
Naphthenes	GC 49		29.7	49.9			36.0
Paraffins	D1319		63.3	8.4			4.8
Isoparaffins			28.6				
Normal Paraffins			34.7				
Aromatics			7.0	41.1			52.1
Olefins	GCM 119	1.22					
BS&W Vol.%	D96	<0.01					
			Arom. Sulphur	0.6			2.9
			Polar Cmpds.	0.0			3.6
			Unidentified	0.0			

TABLE 4.1.2

CRUDE ASSAY, NAPHTHA REFORMING YIELD PREDICTIONS

Predicted Middle of Cycle Yields Based on Feed
(Liquid Volume Percent)

2.000 MPag Pressure
2.000 LHSV

Reformate Octane Number (RON, Clear)	84	85	88	90	92	94	96	98
Benzene	2.0	3.2	3.6	3.9	4.3	4.6	4.9	5.3
Toluene	7.7	8.2	8.8	9.4	10.0	10.5	11.1	11.6
C_8 Alkylbenzenes	10.7	11.3	11.8	12.3	12.8	13.3	13.7	14.2
C_9^+ Alkylbenzenes	4.1	4.2	4.4	4.6	4.8	4.9	5.1	5.3
C_5^+ Reformate Yield	87.4	86.2	85.0	83.7	82.4	81.1	79.8	78.4

Sample D83-12-5-2, Syncrude Naphtha 15-149°C
Project 352-83
Sept. 8, 1983

TABLE 4.1.3 CRUDE ASSAY; MIDDLE DISTILLATE MASS SPECTROSCOPY ANALYSIS

GULF CANADA RESEARCH AND DEVELOPMENT DEPARTMENT

SAMPLE D83-12-5-3 149-385C SYNCR CUT ANALYSED: 1983 SEPT 8. 14:41
PROJECT NO 352-83 RUN NUMBER 181148.5 METHOD DIST

DISTILLATE ANALYSIS BY LOW RESOLUTION MASS SPECTROMETRY

HYDROCARBON TYPE (CN.H2N-Z)	Z NO	DETERMINED VOLUME PERCENT	CALCULATED MASS PERCENT
SATURATES		59.78	58.27
PARAFFINS	+2	9.20	8.39
CYCLOPARAFFINS		50.58	49.88
NONCOND CYCLOPARAFFINS	0	22.58	21.38
COND DICYCLOPARAFFINS	-2	21.24	21.33
COND POLYCYCLOPARAFFINS	-4	6.76	7.18
AROMATICS		40.22	41.73
MONOAROMATICS		32.65	33.14
ALKYLBENZENES	-6	16.65	16.34
BENZOCYCLOPARAFFINS	-8	10.34	10.86
BENZODICYCLOPARAFFINS	-10	5.66	5.94
DIAROMATICS		5.66	6.23
NAPHTHALENES	-12	2.66	2.86
NAPHTHOCYCLOPARAFFINS	-14	1.30	1.44
FLUORENES	-16	1.70	1.94
TRIAROMATICS		1.31	1.56
PHENANTHRENES/ANTHRACENE	-18	1.11	1.30
PHENANTHROCYCLOPARAFFINS	-20	0.21	0.26
TETRAAROMATICS		0.17	0.22
PYRENES/BENZOFLUORENES	-22	0.17	0.22
CHRYSENES/BENZANTHRACENE	-24	0.00	0.00
PENTAAROMATICS		0.00	0.00
CHRYSOCYCLOPARAFFINS	-26	0.00	0.00
BENZPYRENES/PERYLENES	-28	0.00	0.00
DIBENZANTHRACENES	-30	0.00	0.00
UNIDENTIFIED		0.00	0.00
Cn.H2n-32	-32	0.00	0.00
Cn.H2n-36/Cn.H2n-26.S	-36	0.00	0.00
Cn.H2n-38/Cn.H2n-28.S	-38	0.00	0.00
Cn.H2n-42/Cn.H2n-32.S	-42	0.00	0.00
Cn.H2n-44/Cn.H2n-34.S	-44	0.00	0.00
AROMATIC SULFUR		0.43	0.56
BENZOTHIOPHENES	-10S	0.01	0.01
[illegible]THIOPHENES	-16S	0.38	0.50
[illegible]HTHOTHIOPHENES	-22S	0.04	0.05

Table 4.1.4 Crude Assay; Gas Oil Mass Spectroscopy Analysis

GULF CANADA RESEARCH AND DEVELOPMENT DEPARTMENT

SAMPLE D83-12-5-4 SYNCRUDE 385C+ ANALYSED: 1983 SEPT 9. 15:23
PROJECT NO 352-83 RUN NUMBER IB1166.5 METHOD HDSAT

HEAVY DISTILLATE ANALYSIS BY LOW RESOLUTION MASS SPECTROMETRY (GCM 335)

HYDROCARBON TYPE (CN.H2N+Z)	Z NO	MASS %	TOTAL MASS %	GRAND TOTAL MASS %
PARAFFINS	+2		4.85	
CYCLOPARAFFINS			35.97	
MONOCYCLOPARAFFINS	0	9.17		
DICYCLOPARAFFINS	-2	10.15		
TRICYCLOPARAFFINS	-4	7.61		
TETRACYCLOPARAFFINS	-6	6.21		
PENTACYCLOPARAFFINS	-8	2.03		
HEXACYCLOPARAFFINS	-10	0.79		
MONOAROMATICS			19.14	
ALKYLBENZENES	-6	7.53		
BENZOCYCLOPARAFFINS	-8	6.13		
BENZODICYCLOPARAFFINS	-10	5.48		
DIAROMATICS			15.02	
NAPHTHALENES	-12	4.03		
NAPHTHOCYCLOPARAFFINS	-14	4.78		
FLUORENES	-16	6.20		
TRIAROMATICS			9.33	
PHENANTHRENES/ANTHRACENE	-18	5.79		
PHENANTHROCYCLOPARAFFINS	-20	3.53		
TETRAAROMATICS			5.93	
PYRENES/BENZOFLUORENES	-22	3.92		
CHRYSENES/BENZANTHRACENE	-24	2.01		
PENTAAROMATICS			2.67	
CHRYSOCYCLOPARAFFINS	-26	1.53		
BENZPYRENES/PERYLENES	-28	1.06		
DIBENZANTHRACENES	-30	0.08		
UNIDENTIFIED			0.58	
Cn.H2n-32	-32	0.09		
Cn.H2n-36/Cn.H2n-26.S	-36	0.27		
Cn.H2n-38/Cn.H2n-28.S	-38	0.18		
Cn.H2n-42/Cn.H2n-32.S	-42	0.03		
Cn.H2n-44/Cn.H2n-34.S	-44	0.00		
AROMATIC SULFUR			2.92	
BENZOTHIOPHENES	-10S	0.81		
DIBENZOTHIOPHENES	-16S	1.58		
BENZONAPHTHOTHIOPHENES	-22S	0.53		
POLAR COMPOUNDS			3.60	100.00

Table 4.1.5 Crude Assay; Syncrude, G.C. Distillation

REPORT OF GAS CHROMATOGRAPHY DISTILLATION

% OF SAMPLE	BOILING POINT C	BOILING POINT F	% OF SAMPLE	BOILING POINT C	BOILING POINT F	% OF SAMPLE	BOILING POINT C	BOILING POINT F	% OF SAMPLE	BOILING POINT C	BOILING POINT F
IBP	-18	0	25	199	388	50	294	561	75	383	721
1	-9	16	26	203	397	51	297	567	76	387	729
2	5	41	27	207	405	52	300	572	77	391	736
3	27	81	28	212	414	53	303	577	78	395	743
4	38	100	29	216	421	54	309	586	79	400	752
5	56	133	30	220	428	55	312	594	80	404	759
6	68	154	31	225	437	56	316	601	81	409	766
7	73	163	32	229	444	57	320	608	82	412	774
8	83	181	33	233	451	58	323	613	83	417	783
9	91	196	34	237	459	59	327	621	84	422	792
10	99	210	35	241	466	60	330	626	85	427	801
11	105	221	36	245	473	61	333	631	86	432	810
12	116	241	37	248	478	62	337	639	87	437	819
13	121	250	38	251	484	63	340	644	88	443	829
14	130	266	39	255	491	64	343	649	89	448	839
15	138	280	40	258	496	65	347	657	90	454	849
16	144	291	41	262	504	66	351	664	91	462	864
17	151	304	42	266	511	67	354	669	92	470	878
18	159	318	43	270	518	68	359	678	93	477	891
19	164	327	44	273	523	69	362	684	94	485	905
20	170	338	45	277	531	70	365	689	95	494	921
21	174	345	46	280	536	71	369	696	96	505	941
22	181	358	47	283	541	72	372	702	97	518	964
23	187	369	48	287	549	73	376	709	98	534	993
24	194	381	49	290	554	74	379	714	99	557	1035
									FBP	574	1065

TABLE 4.2.1

CATALYST PROPERTIES

CATALYST	NICKEL wt%	MOLYBDENUM wt%	TUNGSTEN wt%	SILICON wt%	ALUMINUM wt%	SIZE	CRUSH STRENGTH lbs	BULK DENSITY g/cc	SURFACE AREA m^2/g	PORE VOLUME cc/g
ARMAK KF840Q	3.2	13.2	0	0	34.7	1.3QLOBE	9	.83	197	0.42
AMER CYAN HDS30	3.9	13.7	0	0	32.5	1/16" E	-	.83	160	0.44
KATALCO NM506	5.3	18.0	0	0	29.5	1/16" E	14	.92	220	0.39
KATALCO NT550	3.9	0	15.9	0	34.9	1/16" E	16	.81	230	0.50
SHELL S324	2.7	13.2	0	0	32.6	1/16" E	18	.86	134	0.48
GULF CATALYST A*	- - - Proprietary - - -			mixed SiO_2 - Al_2O_3		1/12" E	12	.87	203	-
SHELL 454	5	0	23	0.3	30.1	1/16" E	18	.96	150	0.3
HARSHAW 4301E	5.1	0	19	25.5	4.6	1/16" E	17	.96	230	0.37

* Refer to U.S. Patent No. 4427534, January 24, 1984.

TABLE 4.2.2

CATALYST SCREENING

Feedstock: Syncrude Middle Distillate 149-385°C
Catalyst : American Cyanamid HDN30

RUN #	6031	6029	6030
Run Conditions			
Pressure, MPag(psig)	13.8 (2000)	13.8 (2000)	13.8 (2000)
W.A.B.Temperature, °C (°F)	340 (644)	385 (725)	420 (788)
L.H.S.V.	1.0	1.0	1.0
Gas Rate, L/L (SCFB)	890 (5000)	890 (5000)	890 (5000)
H_2 Consumption, L/L (SCFB)	52 (294)	121 (681)	137 (771)
Yields			
Product sg @ 15°C	0.8737	0.8487	0.8260
Feed sg @ 15°C	0.8838	0.8838	0.8838
Vol./vol. % yield of TLP on Feed	101.2	104.1	107.0
Product Inspections			
GCD Light Ends, liq. vol. %			
Ethane	nil	0.03	0.06
Propane	0.02	0.20	0.56
IsoButane	nil	0.10	0.34
N-Butane	0.05	0.39	1.03
IsoPentane	nil	0.19	0.64
N-Pentane	0.04	0.33	0.90
C_6+	99.89	98.77	96.45
Distillation, D86 °C (°F)			
IBP	164 (327)	78 (172)	74 (165)
10%	208 (406)	180 (356)	142 (288)
20%	225 (437)	203 (397)	172 (342)
30%	244 (471)	225 (437)	195 (383)
40%	261 (502)	241 (466)	211 (412)
50%	275 (527)	257 (495)	228 (442)
60%	291 (556)	273 (523)	244 (471)
70%	305 (581)	289 (552)	261 (502)
80%	319 (606)	306 (583)	284 (543)
90%	337 (639)	329 (624)	311 (592)
FBP	349 (660)	343 (649)	326 (619)
Estimated Molecular Weight	201	189	169
Sulphur, wt ppm	67	34	16
Nitrogen, wt ppm	<1	<1	<1
Naphtha Make, wt% on Feed	nil	1.2	5.2
Product Composition, Mass Spec. wt% (149-385°C product only)			
Paraffins	10.9	11.8	13.2
Cycloparaffins	51.2	71.5	70.1
Monoaromatics	36.1	16.0	16.5
Diaromatics	0.4	0.2	0.2
Triaromatics+	0.6	0.5	0.0
Thiophenes	0.8	0.0	0.0

TABLE 4.2.3

CATALYST SCREENING

Feedstock: Syncrude Middle Distillate 149-385°C
Catalyst : HARSHAW NI4301E

RUN #	6051	6052	6053
Run Conditions			
Pressure, MPag (psig)	13.8 (2000)	13.8 (2000)	13.8 (2000)
W.A.B.Temperature, °C (°F)	340 (644)	385 (725)	420 (788)
L.H.S.V.	1.0	1.0	1.0
Gas Rate, L/L (SCFB)	890 (5000)	890 (5000)	890 (5000)
H_2 Consumption, L/L (SCFB)	54 (306)	157 (881)	165 (928)
Yields			
Product sg @ 15°C	0.8680	0.7740	0.7640
Feed sg @ 15°C	0.8838	0.8838	0.8838
Vol./vol. % yield of TLP on Feed	101.8	110.5	111.2
Product Inspections			
GCD Light Ends, liq. vol. %			
Ethane	nil	0.08	0.15
Propane	0.11	1.41	1.60
IsoButane	0.07	2.78	2.42
N-Butane	0.26	3.35	3.45
IsoPentane	0.13	5.24	4.94
N-Pentane	0.19	2.19	2.77
C_6+	99.24	84.96	84.67
Distillation, D86 °C (°F)			
IBP	113 (235)	32 (90)	34 (93)
10%	193 (379)	73 (163)	72 (162)
20%	218 (424)	98 (208)	92 (198)
30%	237 (459)	118 (244)	108 (226)
40%	254 (489)	136 (277)	122 (252)
50%	271 (520)	154 (309)	137 (279)
60%	287 (549)	173 (343)	152 (306)
70%	303 (577)	192 (378)	169 (336)
80%	319 (606)	218 (424)	190 (374)
90%	339 (642)	275 (527)	235 (455)
FBP	364 (687)	290 (554)	262 (504)
Estimated Molecular Weight	196	156	131
Sulphur, wt ppm	38	5	<1
Nitrogen, wt ppm	4	<1	<1
Naphtha Make, vol% on Feed	4.0	47.5	56.2
Product Composition, Mass Spec. wt%			
Paraffins	11.2	15.1	22.9
Cycloparaffins	56.0	80.3	75.4
Monoaromatics	30.5	4.6	1.6
Diaromatics	0.3	0.1	0.0
Triaromatics+	1.3	0.0	0.0
Thiophenes	0.6	0.0	0.0

TABLE 4.2.4

CATALYST SCREENING

Feedstock: Syncrude Middle Distillate 149-385°C
Catalyst : Katalco NM506

RUN #	6006	6007	6008
Run Conditions			
Pressure, MPag(psig)	13.8 (2000)	13.8 (2000)	13.8 (2000)
W.A.B.Temperature, °C (°F)	340 (644)	385 (725)	420 (788)
L.H.S.V.	1.0	1.0	1.0
Gas Rate, L/L (SCFB)	890 (5000)	890 (5000)	890 (5000)
H_2 Consumption, L/L (SCFB)	132 (740)	151 (850)	159 (892)
Yields			
Product sg @ 15°C	0.8519	0.8395	0.8182
Feed sg @ 15°C	0.8838	0.8838	0.8838
Vol./vol. % yield of TLP on Feed	103.7	105.2	108.0
Product Inspections			
GCD Light Ends, liq. vol. %			
Ethane	nil	0.03	0.12
Propane	0.07	0.22	0.70
IsoButane	0.02	0.15	0.40
N-Butane	0.19	0.41	1.16
IsoPentane	0.06	0.22	0.70
N-Pentane	0.16	0.33	1.00
C_6^+	99.51	98.64	95.93
Distillation, D86 °C (°F)			
IBP	130 (266)	80 (176)	63 (145)
10%	199 (390)	181 (358)	147 (297)
20%	215 (419)	203 (397)	176 (349)
30%	235 (455)	223 (433)	197 (387)
40%	252 (486)	240 (464)	214 (417)
50%	267 (513)	260 (500)	229 (444)
60%	283 (541)	271 (520)	243 (469)
70%	297 (567)	286 (547)	260 (500)
80%	312 (594)	303 (577)	277 (531)
90%	331 (628)	324 (615)	304 (579)
FBP	347 (657)	341 (646)	323 (613)
Estimated Molecular Weight	198	190	172
Sulphur, wt ppm	60	24	2
Nitrogen, wt ppm	4	2	<1
Naphtha Make, wt% on Feed	nil	1.1	7.0
Product Composition, Mass Spec. wt% (149-385°C product only)			
Paraffins	11.0	11.5	12.8
Cycloparaffins	79.9	82.7	77.0
Monoaromatics	7.6	5.1	10.2
Diaromatics	0.1	0.1	0.0
Triaromatics+	1.2	0.7	0.0
Thiophenes	0.2	0.0	0.0

TABLE 4.2.5

CATALYST SCREENING

Feedstock: Syncrude Middle Distillate 149-385°C
Catalyst : Katalco MT550

RUN #	6001	6002	6003
Run Conditions			
Pressure, MPag(psig)	13.8 (2000)	13.8 (2000)	13.8 (2000)
W.A.B.Temperature, °C (°F)	340 (644)	380 (716)	420 (788)
L.H.S.V.	1.0	1.0	1.0
Gas Rate, L/L (SCFB)	890 (5000)	890 (5000)	890 (5000)
H_2 Consumption, L/L (SCFB)	114 (640)	146 (817)	139 (780)
Yields			
Product sg @ 15°C	0.8575	0.8461	0.8306
Feed sg @ 15°C	0.8838	0.8838	0.8838
Vol./vol. % yield of TLP on Feed	103.1	104.5	106.5
Product Inspections			
GCD Light Ends, liq. vol. %			
Ethane	nil	0.03	0.09
Propane	0.10	0.13	0.54
IsoButane	0.02	0.06	0.28
N-Butane	0.14	0.25	0.92
IsoPentane	0.06	0.11	0.48
N-Pentane	0.10	0.23	0.80
C_6^+	99.57	99.19	96.88
Distillation, D86 °C (°F)			
IBP	152 (306)	103 (217)	68 (154)
10%	192 (192)	186 (367)	160 (320)
20%	218 (424)	208 (406)	187 (367)
30%	235 (455)	226 (439)	208 (406)
40%	252 (486)	242 (468)	221 (430)
50%	267 (513)	256 (493)	239 (462)
60%	282 (540)	272 (522)	255 (491)
70%	297 (567)	288 (550)	270 (518)
80%	312 (594)	305 (581)	288 (550)
90%	330 (626)	324 (615)	312 (594)
FBP	346 (655)	340 (644)	329 (624)
Estimated Molecular Weight	195	191	178
Sulphur, wt ppm	1	67	3
Nitrogen, wt ppm	1	4	3
Naphtha Make, wt% on Feed	nil	1.3	4.7
Product Composition, Mass Spec. wt% (149-385°C product only)			
Paraffins	11.4	11.2	11.6
Cycloparaffins	73.6	82.6	74.5
Monoaromatics	13.4	5.3	13.5
Diaromatics	0.0	0.1	0.4
Triaromatics+	1.1	0.8	0.0
Thiophenes	0.5	0.1	0.0

TABLE 4.2.6

CATALYST SCREENING

Feedstock: Syncrude Middle Distillate 149-385°C
Catalyst : Arnak KF840C

RUN #	6009	6010	6011
Run Conditions			
Pressure, MPag (psig)	13.8 (2000)	13.8 (2000)	13.8 (2000)
W.A.B. Temperature, °C (°F)	340 (644)	385 (725)	420 (788)
L.H.S.V.	1.0	1.0	1.0
Gas Rate, L/L (SCFB)	890 (5000)	890 (5000)	890 (5000)
H_2 Consumption, L/L (SCFB)	107 (602)	136 (765)	146 (821)
Yields			
Product sg @ 15°C	0.8585	0.8442	0.8191
Feed sg @ 15°C	0.8838	0.8838	0.8838
Vol./vol. % yield of TLP on Feed	102.9	104.6	107.8
Product Inspections			
GCD Light Ends, liq. vol. %			
Ethane	nil	0.05	0.13
Propane	0.06	0.20	0.77
IsoButane	0.02	0.07	0.38
N-Butane	0.14	0.39	1.23
IsoPentane	0.05	0.20	0.78
N-Pentane	0.14	0.26	1.06
C_6+	99.59	98.83	95.65
Distillation, D86 °C (°F)			
IBP	149 (300)	69 (156)	59 (138)
10%	203 (397)	179 (354)	145 (293)
20%	219 (426)	201 (394)	175 (347)
30%	234 (453)	222 (432)	196 (385)
40%	255 (491)	239 (462)	215 (419)
50%	270 (518)	256 (493)	229 (444)
60%	283 (541)	270 (518)	244 (471)
70%	300 (572)	286 (547)	261 (502)
80%	313 (595)	261 (502)	279 (534)
90%	332 (630)	323 (613)	304 (579)
FBP	350 (662)	344 (651)	328 (622)
Estimated Molecular Weight	199	188	171
Sulphur, wt ppm	1	1	1
Nitrogen, wt ppm	<1	<1	<1
Naphtha Make, wt% on Feed	nil	nil	7.0
Product Composition, Mass Spec. wt% (149-385°C product only)			
Paraffins	11.8	12.6	12.3
Cycloparaffins	69.1	75.6	74.4
Monoaromatics	17.9	11.1	12.8
Diaromatics	0.0	0.1	0.4
Triaromatics+	1.0	0.6	0.1
Thiophenes	0.3	0.0	0.0

TABLE 4.2.7

CATALYST SCREENING

Feedstock: Syncrude Middle Distillate 149-385°C
Catalyst : Shell S324

RUN #	6018	6019	6020
Run Conditions			
Pressure, MPag (psig)	13.8 (2000)	13.8 (2000)	13.8 (2000)
W.A.B.Temperature, °C (°F)	340 (644)	385 (725)	420 (788)
L.H.S.V.	1.0	1.0	1.0
Gas Rate, L/L (SCFB)	890 (5000)	890 (5000)	890 (5000)
H_2 Consumption, L/L (SCFB)	83 (468)	132 (742)	139 (781)
Yields			
Product sg @ 15°C	0.8643	0.8448	0.8269
Feed sg @ 15°C	0.8838	0.8838	0.8838
Vol./vol. % yield of TLP on Feed	102.2	104.6	106.8
Product Inspections			
GCD Light Ends, liq. vol. %			
Ethane	nil	0.03	0.07
Propane	0.05	0.19	0.60
IsoButane	0.02	0.16	0.35
N-Butane	0.09	0.42	1.05
IsoPentane	0.04	0.20	0.56
N-Pentane	0.07	0.34	0.91
C_6+	99.73	98.66	96.46
Distillation, D86 °C (°F)			
IBP	155 (311)	93 (199)	60 (140)
10%	197 (387)	178 (352)	152 (306)
20%	215 (419)	201 (394)	182 (360)
30%	235 (455)	221 (430)	203 (397)
40%	250 (482)	238 (460)	220 (428)
50%	266 (511)	253 (487)	236 (457)
60%	280 (536	268 (514)	252 (486)
70%	296 (565)	284 (543)	268 (514)
80%	312 (594)	302 (576)	286 (547)
90%	332 (630)	324 (615)	313 (595)
FBP	348 (658)	340 (644)	330 (626)
Estimated Molecular Weight	195	187	175
Sulphur, wt ppm	8	3	2
Nitrogen, wt ppm	<1	1	1
Naphtha Make, wt% on Feed	nil	2.8	6.8
Product Composition, Mass Spec. wt% (149-385°C product only)			
Paraffins	11.7	11.8	12.4
Cycloparaffins	60.1	75.4	72.6
Monoaromatics	27.0	12.1	14.8
Diaromatics	0.1	0.1	0.2
Triaromatics+	0.4	0.5	0.0
Thiophenes	0.7	0.0	0.0

TABLE 4.2.8

CATALYST SCREENING

Feedstock: Syncrude Middle Distillate 149-385°C
Catalyst : SHELL S454

RUN #	6048	6049	6050
Run Conditions			
Pressure, MPag(psig)	13.8 (2000)	13.8 (2000)	13.8 (2000)
W.A.B.Temperature, °C (°F)	340 (644)	385 (725)	420 (788)
L.H.S.V.	1.0	1.0	1.0
Gas Rate, L/L (SCFB)	890 (5000)	890 (5000)	890 (5000)
H_2 Consumption, L/L (SCFB)	133 (748)	137 (772)	168 (943)
Yields			
Product sg @ 15°C	0.8588	0.8510	0.8240
Feed sg @ 15°C	0.8838	0.8838	0.8838
Vol./vol. % yield of TLP on Feed	102.9	103.9	107.3
Product Inspections			
GCD Light Ends, liq. vol. %			
Ethane	nil	0.05	0.12
Propane	0.01	0.25	0.76
IsoButane	nil	0.09	0.33
N-Butane	0.03	0.32	1.14
IsoPentane	nil	0.14	0.56
N-Pentane	0.03	0.23	0.90
C_6+	99.94	98.93	96.19
Distillation, D86 °C (°F)			
IBP	174 (345)	112 (234)	52 (126)
10%	207 (405)	191 (376)	153 (307)
20%	225 (437)	215 (419)	182 (360)
30%	243 (469)	233 (451)	202 (396)
40%	259 (498)	249 (480)	219 (426)
50%	274 (525)	266 (511)	234 (453)
60%	289 (552)	281 (538)	249 (480)
70%	303 (577)	295 (563)	264 (507)
80%	318 (604)	312 (594)	282 (540)
90%	338 (640)	332 (630)	307 (585)
FBP	357 (675)	354 (669)	332 (630)
Estimated Molecular Weight	204	195	174
Sulphur, wt ppm	48	23	5
Nitrogen, wt ppm	15	7	4
Naphtha Make, vol% on Feed	nil	3.8	10.2
Product Composition, Mass Spec. wt% (149-385°C product only)			
Paraffins	10.0	12.6	10.7
Cycloparaffins	80.5	79.3	79.3
Monoaromatics	7.2	7.9	8.0
Diaromatics	0.1	0.2	0.4
Triaromatics+	1.6	0.0	1.3
Thiophenes	0.6	0.0	0.3

TABLE 4.2.9

CATALYST SCREENING

Feedstock: Syncrude Middle Distillate 149-385°C
Catalyst : GULF CATALYST A

RUN #	6012	6014	6013
Run Conditions			
Pressure, MPag(psig)	13.8 (2000)	13.8 (2000)	13.8 (2000)
W.A.B.Temperature, °C (°F)	340 (644)	360 (680)	385 (725)
L.H.S.V.	1.0	1.0	1.0
Gas Rate, L/L (SCFB)	890 (5000)	890 (5000)	890 (5000)
H_2 Consumption, L/L (SCFB)	115 (646)	140 (784)	187 (1049)
Yields			
Product sg @ 15°C	0.8533	0.8414	0.8037
Feed sg @ 15°C	0.8838	0.8838	0.8838
Vol./vol. % yield of TLP on Feed	103.5	105.0	109.9
Product Inspections			
GCD Light Ends, liq. vol. %			
Ethane	nil	nil	0.03
Propane	0.11	0.17	0.61
IsoButane	0.06	0.17	1.05
N-Butane	0.25	0.41	1.26
IsoPentane	0.12	0.27	1.57
N-Pentane	0.22	0.31	1.11
C_6^+	99.23	98.68	94.35
Distillation, D86 °C (°F)			
IBP	124 (255)	110 (230)	61 (142)
10%	184 (363)	168 (334)	121 (250)
20%	210 (410)	197 (387)	151 (304)
30%	230 (446)	216 (421)	174 (345)
40%	247 (477)	235 (455)	193 (379)
50%	265 (509)	250 (482)	209 (408)
60%	280 (536)	268 (514)	226 (439)
70%	295 (563)	285 (545)	243 (469)
80%	312 (594)	304 (579)	262 (504)
90%	332 (630)	326 (619)	294 (561)
FBP	347 (657)	342 (648)	316 (601)
Estimated Molecular Weight	193	181	159
Sulphur, wt ppm	12	10	1
Nitrogen, wt ppm	1	1	1
Naphtha Make, wt% on Feed	nil	3.3	11.5
Product Composition, Mass Spec. wt% (149-385°C product only)			
Paraffins	11.8	11.3	13.1
Cycloparaffins	71.0	79.4	86.1
Monoaromatics	16.1	8.6	0.8
Diaromatics	0.2	0.1	0.0
Triaromatics+	0.9	0.7	0.0
Thiophenes	0.0	0.0	0.0

GULF CANADA RESEARCH AND DEVELOPMENT DEPARTMENT

TABLE 4.2.10 MASS SPECTROSCOPY DATA COMPARISON : AMERICAN CYANAMID HDN30

FEED

RUN #	6031		6029		6030	
HYDROCARBON TYPE	CALCULATED MASS PERCENT					
SATURATES		55.24		62.13		83.31
PARAFFINS	11.20		10.89		13.22	
CYCLOPARAFFINS		44.04		51.23		70.09
NONCOND CYCLOPARAFFINS	19.98		22.14		36.36	
COND DICYCLOPARAFFINS	16.63		20.37		27.82	
COND POLYCYCLOPARAFFINS	7.43		8.72		5.91	
AROMATICS		44.76		37.87		16.69
MONOAROMATICS		35.05		36.09		16.47
ALKYLBENZENES	17.11		16.52		9.69	
BENZOCYCLOPARAFFINS	11.90		11.99		4.98	
BENZODICYCLOPARAFFINS	6.04		7.58		1.79	
DIAROMATICS		6.99		0.43		0.22
NAPHTHALENES	3.09		0.43		0.22	
NAPHTHOCYCLOPARAFFINS	1.66		0.00		0.00	
FLUORENES	2.24		0.00		0.00	
TRIAROMATICS		1.74		0.53		0.00
PHENANTHRENES/ANTHRACENE	1.42		0.44		0.00	
PHENANTHROCYCLOPARAFFINS	0.32		0.08		0.00	
TETRAAROMATICS		0.21		0.06		0.00
PYRENES/BENZOFLUORENES	0.19		0.06		0.00	
CHRYSENES/BENZANTHRACENE	0.02		0.00		0.00	
PENTAAROMATICS		0.00		0.00		0.00
CHRYSOCYCLOPARAFFINS	0.00		0.00		0.00	
BENZPYRENES/PERYLENES	0.00		0.00		0.00	
DIBENZANTHRACENES	0.00		0.00		0.00	
UNIDENTIFIED		0.00		0.00		0.00
Cn.H2n-32	0.00		0.00		0.00	
Cn.H2n-36/Cn.H2n-26.S	0.00		0.00		0.00	
Cn.H2n-38/Cn.H2n-28.S	0.00		0.00		0.00	
Cn.H2n-42/Cn.H2n-32.S	0.00		0.00		0.00	
Cn.H2n-44/Cn.H2n-34.S	0.00		0.00		0.00	
AROMATIC SULFUR		0.77		0.77		0.00
BENZOTHIOPHENES	0.07		0.00		0.00	
DIBENZOTHIOPHENES	0.69		0.77		0.00	
BENZONAPHTHOTHIOPHENES	0.02		0.00		0.00	

GULF CANADA RESEARCH AND DEVELOPMENT DEPARTMENT

TABLE 4.2.11 MASS SPECTROSCOPY DATA COMPARISON : HARSHAW N14301E

	FEED							
RUN #		6051		6052		6053		
HYDROCARBON TYPE	CALCULATED MASS PERCENT							
SATURATES		55.24		67.23		95.39		98.36
PARAFFINS	11.20		11.21		15.12		22.94	
CYCLOPARAFFINS		44.04		56.03		80.26		75.43
NONCOND CYCLOPARAFFINS	19.98		24.90		46.72		58.12	
COND DICYCLOPARAFFINS	16.63		21.63		27.95		16.31	
COND POLYCYCLOPARAFFINS	7.43		9.50		5.59		0.99	
AROMATICS		44.76		32.77		4.61		1.64
MONOAROMATICS		35.05		30.51		4.55		1.64
ALKYLBENZENES	17.11		15.37		4.14		1.38	
BENZOCYCLOPARAFFINS	11.90		9.37		0.39		0.26	
BENZODICYCLOPARAFFINS	6.04		5.76		0.02		0.00	
DIAROMATICS		6.99		0.33		0.06		0.00
NAPHTHALENES	3.09		0.33		0.06		0.00	
NAPHTHOCYCLOPARAFFINS	1.66		0.00		0.00		0.00	
FLUORENES	2.24		0.00		0.00		0.00	
TRIAROMATICS		1.74		0.47		0.00		0.00
PHENANTHRENES/ANTHRACENE	1.42		0.47		0.00		0.00	
PHENANTHROCYCLOPARAFFINS	0.32		0.00		0.00		0.00	
TETRAAROMATICS		0.21		0.04		0.00		0.00
PYRENES/BENZOFLUORENES	0.19		0.04		0.00		0.00	
CHRYSENES/BENZANTHRACENE	0.02		0.00		0.00		0.00	
PENTAAROMATICS		0.00		0.82		0.00		0.00
CHRYSOCYCLOPARAFFINS	0.00		0.00		0.00		0.00	
BENZPYRENES/PERYLENES	0.00		0.00		0.00		0.00	
DIBENZANTHRACENES	0.00		0.82		0.00		0.00	
UNIDENTIFIED		0.00		0.00		0.00		0.00
Cn.H2n-32	0.00		0.00		0.00		0.00	
Cn.H2n-36/Cn.H2n-26.8	0.00		0.00		0.00		0.00	
Cn.H2n-38/Cn.H2n-28.8	0.00		0.00		0.00		0.00	
Cn.H2n-42/Cn.H2n-32.8	0.00		0.00		0.00		0.00	
Cn.H2n-44/Cn.H2n-34.8	0.00		0.00		0.00		0.00	
AROMATIC SULFUR		0.77		0.60		0.00		0.00
BENZOTHIOPHENES	0.07		0.00		0.00		0.00	
DIBENZOTHIOPHENES	0.69		0.60		0.00		0.00	
BENZONAPHTHOTHIOPHENES	0.02		0.00		0.00		0.00	

GULF CANADA RESEARCH AND DEVELOPMENT DEPARTMENT

TABLE 4.2.12 MASS SPECTROSCOPY DATA COMPARISON : KATALCO NM506

	FEED	6006	6007	6008
RUN #				
HYDROCARBON TYPE	CALCULATED MASS PERCENT			
SATURATES	55.24	90.88	94.19	89.79
PARAFFINS	11.20	10.98	11.50	12.84
CYCLOPARAFFINS	44.04	79.90	82.70	76.95
NONCOND CYCLOPARAFFINS	19.98	33.90	38.31	43.65
COND DICYCLOPARAFFINS	16.63	31.65	33.14	29.06
COND POLYCYCLOPARAFFINS	7.43	14.35	11.25	4.23
AROMATICS	44.76	9.12	5.81	10.21
MONOAROMATICS	35.05	7.64	5.07	10.21
ALKYLBENZENES	17.11	5.50	3.33	5.10
BENZOCYCLOPARAFFINS	11.90	1.44	1.06	3.94
BENZODICYCLOPARAFFINS	6.04	0.69	0.67	1.17
DIAROMATICS	6.99	0.09	0.09	0.00
NAPHTHALENES	3.09	0.09	0.09	0.00
NAPHTHOCYCLOPARAFFINS	1.66	0.00	0.00	0.00
FLUORENES	2.24	0.00	0.00	0.00
TRIAROMATICS	1.74	1.12	0.61	0.00
PHENANTHRENES/ANTHRACENE	1.42	0.83	0.56	0.00
PHENANTHROCYCLOPARAFFINS	0.32	0.30	0.05	0.00
TETRAAROMATICS	0.21	0.11	0.04	0.00
PYRENES/BENZOFLUORENES	0.19	0.11	0.04	0.00
CHRYSENES/BENZANTHRACENE	0.02	0.00	0.00	0.00
PENTAAROMATICS	0.00	0.00	0.00	0.00
CHRYSOCYCLOPARAFFINS	0.00	0.00	0.00	0.00
BENZPYRENES/PERYLENES	0.00	0.00	0.00	0.00
DIBENZANTHRACENES	0.00	0.00	0.00	0.00
UNIDENTIFIED	0.00	0.00	0.00	0.00
Cn.H2n-32	0.00	0.00	0.00	0.00
Cn.H2n-36/Cn.H2n-26.S	0.00	0.00	0.00	0.00
Cn.H2n-38/Cn.H2n-28.S	0.00	0.00	0.00	0.00
Cn.H2n-42/Cn.H2n-32.S	0.00	0.00	0.00	0.00
Cn.H2n-44/Cn.H2n-34.S	0.00	0.00	0.00	0.00
AROMATIC SULFUR	0.77	0.16	0.00	0.00
BENZOTHIOPHENES	0.07	0.16	0.00	0.00
DIBENZOTHIOPHENES	0.69	0.00	0.00	0.00
BENZONAPHTHOTHIOPHENES	0.02	0.00	0.00	0.00

GULF CANADA RESEARCH AND DEVELOPMENT DEPARTMENT

TABLE 4.2.13 MASS SPECTROSCOPY DATA COMPARISON : KATALCO NT550

	FEED	6001		6002		6003	
RUN #							
	CALCULATED MASS PERCENT						
HYDROCARBON TYPE							
SATURATES	55.24		84.95		93.82		86.09
PARAFFINS	11.20	11.37		11.22		11.56	
CYCLOPARAFFINS	44.04		73.59		82.60		74.53
NONCOND CYCLOPARAFFINS	19.98	30.43		36.84		39.18	
COND DICYCLOPARAFFINS	16.63	29.63		32.95		28.54	
COND POLYCYCLOPARAFFINS	7.43	13.52		12.81		6.81	
AROMATICS	44.76		15.05		6.18		13.91
MONOAROMATICS	35.05		13.43		5.25		13.54
ALKYLBENZENES	17.11	7.56		2.98		5.55	
BENZOCYCLOPARAFFINS	11.90	3.69		1.31		5.48	
BENZODICYCLOPARAFFINS	6.04	2.18		0.96		2.51	
DIAROMATICS	6.99		0.00		0.10		0.35
NAPHTHALENES	3.09	0.00		0.10		0.35	
NAPHTHOCYCLOPARAFFINS	1.66	0.00		0.00		0.00	
FLUORENES	2.24	0.00		0.00		0.00	
TRIAROMATICS	1.74		1.05		0.74		0.00
PHENANTHRENES/ANTHRACENE	1.42	0.80		0.59		0.00	
PHENANTHROCYCLOPARAFFINS	0.32	0.24		0.14		0.00	
TETRAAROMATICS	0.21		0.03		0.02		0.01
PYRENES/BENZOFLUORENES	0.19	0.03		0.02		0.01	
CHRYSENES/BENZANTHRACENE	0.02	0.00		0.00		0.00	
PENTAAROMATICS	0.00		0.00		0.00		0.00
CHRYSOCYCLOPARAFFINS	0.00	0.00		0.00		0.00	
BENZPYRENES/PERYLENES	0.00	0.00		0.00		0.00	
DIBENZANTHRACENES	0.00	0.00		0.00		0.00	
UNIDENTIFIED	0.00		0.00		0.00		0.00
Cn.H2n-32	0.00	0.00		0.00		0.00	
Cn.H2n-36/Cn.H2n-26.S	0.00	0.00		0.00		0.00	
Cn.H2n-38/Cn.H2n-28.S	0.00	0.00		0.00		0.00	
Cn.H2n-42/Cn.H2n-32.S	0.00	0.00		0.00		0.00	
Cn.H2n-44/Cn.H2n-34.S	0.00	0.00		0.00		0.00	
AROMATIC SULFUR	0.77		0.54		0.07		0.00
BENZOTHIOPHENES	0.07	0.22		0.07		0.00	
DIBENZOTHIOPHENES	0.69	0.33		0.00		0.00	
BENZONAPHTHOTHIOPHENES	0.02	0.00		0.00		0.00	

GULF CANADA RESEARCH AND DEVELOPMENT DEPARTMENT

TABLE 4.2.14 MASS SPECTROSCOPY DATA COMPARISON : KETJEN ARMAK KF840Q

	FEED			
RUN #		6009	6010	6011
HYDROCARBON TYPE	CALCULATED MASS PERCENT			
SATURATES	55.24	80.85	88.19	86.69
PARAFFINS	11.20	11.75	12.60	12.27
CYCLOPARAFFINS	44.04	69.10	75.59	74.42
NONCOND CYCLOPARAFFINS	19.98	28.32	33.16	39.90
COND DICYCLOPARAFFINS	16.63	27.79	31.04	29.47
COND POLYCYCLOPARAFFINS	7.43	12.99	11.40	5.05
AROMATICS	44.76	19.15	11.81	13.31
MONOAROMATICS	35.05	17.92	11.08	12.84
ALKYLBENZENES	17.11	10.95	7.66	7.79
BENZOCYCLOPARAFFINS	11.90	4.81	2.31	3.82
BENZODICYCLOPARAFFINS	6.04	2.16	1.10	1.23
DIAROMATICS	6.99	0.00	0.11	0.38
NAPHTHALENES	3.09	0.00	0.11	0.38
NAPHTHOCYCLOPARAFFINS	1.66	0.00	0.00	0.00
FLUORENES	2.24	0.00	0.00	0.00
TRIAROMATICS	1.74	0.78	0.61	0.01
PHENANTHRENES/ANTHRACENE	1.42	0.64	0.55	0.00
PHENANTHROCYCLOPARAFFINS	0.32	0.14	0.06	0.01
TETRAAROMATICS	0.21	0.18	0.01	0.07
PYRENES/BENZOFLUORENES	0.19	0.11	0.01	0.07
CHRYSENES/BENZANTHRACENE	0.02	0.08	0.00	0.00
PENTAAROMATICS	0.00	0.00	0.00	0.00
CHRYSOCYCLOPARAFFINS	0.00	0.00	0.00	0.00
BENZPYRENES/PERYLENES	0.00	0.00	0.00	0.00
DIBENZANTHRACENES	0.00	0.00	0.00	0.00
UNIDENTIFIED	0.00	0.00	0.00	0.00
Cn.H2n-32	0.00	0.00	0.00	0.00
Cn.H2n-36/Cn.H2n-26.8	0.00	0.00	0.00	0.00
Cn.H2n-38/Cn.H2n-28.8	0.00	0.00	0.00	0.00
Cn.H2n-42/Cn.H2n-32.8	0.00	0.00	0.00	0.00
Cn.H2n-44/Cn.H2n-34.8	0.00	0.00	0.00	0.00
AROMATIC SULFUR	0.77	0.26	0.00	0.00
BENZOTHIOPHENES	0.07	0.04	0.00	0.00
DIBENZOTHIOPHENES	0.69	0.22	0.00	0.00
BENZONAPHTHOTHIOPHENES	0.02	0.00	0.00	0.00

GULF CANADA RESEARCH AND DEVELOPMENT DEPARTMENT

TABLE 4.2.15 MASS SPECTROSCOPY DATA COMPARISON : SHELL S324

	FEED	RUN # 6018		6019		6020	
HYDROCARBON TYPE	CALCULATED MASS PERCENT						
SATURATES	55.24		71.79		87.22		85.03
PARAFFINS	11.20	11.70		11.81		12.43	
CYCLOPARAFFINS	44.04		60.09		75.41		72.59
NONCOND CYCLOPARAFFINS	19.98	25.18		32.68		37.03	
COND DICYCLOPARAFFINS	16.63	24.27		31.16		28.93	
COND POLYCYCLOPARAFFINS	7.43	10.64		11.57		6.63	
AROMATICS	44.76		28.21		12.78		14.97
MONOAROMATICS	35.05	27.02		12.14		14.82	
ALKYLBENZENES	17.11	13.56		7.71		8.20	
BENZOCYCLOPARAFFINS	11.90	8.64		2.96		4.70	
BENZODICYCLOPARAFFINS	6.04	4.82		1.47		1.93	
DIAROMATICS	6.99		0.10		0.11		0.15
NAPHTHALENES	3.09	0.10		0.11		0.15	
NAPHTHOCYCLOPARAFFINS	1.66	0.00		0.00		0.00	
FLUORENES	2.24	0.00		0.00		0.00	
TRIAROMATICS	1.74		0.39		0.51		0.00
PHENANTHRENES/ANTHRACENE	1.42	0.38		0.46		0.00	
PHENANTHROCYCLOPARAFFINS	0.32	0.02		0.05		0.00	
TETRAAROMATICS	0.21		0.01		0.01		0.00
PYRENES/BENZOFLUORENES	0.19	0.01		0.01		0.00	
CHRYSENES/BENZANTHRACENE	0.02	0.00		0.00		0.00	
PENTAAROMATICS	0.00		0.00		0.00		0.00
CHRYSOCYCLOPARAFFINS	0.00	0.00		0.00		0.00	
BENZPYRENES/PERYLENES	0.00	0.00		0.00		0.00	
DIBENZANTHRACENES	0.00	0.00		0.00		0.00	
UNIDENTIFIED	0.00		0.00		0.00		0.00
Cn.H2n-32	0.00	0.00		0.00		0.00	
Cn.H2n-36/Cn.H2n-26.8	0.00	0.00		0.00		0.00	
Cn.H2n-38/Cn.H2n-28.8	0.00	0.00		0.00		0.00	
Cn.H2n-42/Cn.H2n-32.8	0.00	0.00		0.00		0.00	
Cn.H2n-44/Cn.H2n-34.8	0.00	0.00		0.00		0.00	
AROMATIC SULFUR	0.77		0.69		0.00		0.00
BENZOTHIOPHENES	0.07	0.00		0.00		0.00	
DIBENZOTHIOPHENES	0.69	0.69		0.00		0.00	
BENZONAPHTHOTHIOPHENES	0.02	0.00		0.00		0.00	

GULF CANADA RESEARCH AND DEVELOPMENT DEPARTMENT

TABLE 4.2.16 MASS SPECTROSCOPY DATA COMPARISON : SHELL S454

	FEED							
	RUN #	6048		6049		6050		
HYDROCARBON TYPE	CALCULATED MASS PERCENT							
SATURATES	55.24		90.52		89.97		91.86	
PARAFFINS	11.20	10.03		10.70		12.59		
CYCLOPARAFFINS	44.04		80.49		79.27		79.26	
NONCOND CYCLOPARAFFINS	19.98	34.87		37.28		43.90		
COND DICYCLOPARAFFINS	16.63	32.94		31.36		30.56		
COND POLYCYCLOPARAFFINS	7.43	12.68		10.63		4.80		
AROMATICS	44.76		9.48		10.03		8.14	
MONOAROMATICS	35.05		7.18		7.97		7.89	
ALKYLBENZENES	17.11	3.67		2.86		2.42		
BENZOCYCLOPARAFFINS	11.90	1.98		2.83		3.76		
BENZODICYCLOPARAFFINS	6.04	1.53		2.28		1.72		
DIAROMATICS	6.99		0.13		0.40		0.20	
NAPHTHALENES	3.09	0.13		0.40		0.20		
NAPHTHOCYCLOPARAFFINS	1.66	0.00		0.00		0.00		
FLUORENES	2.24	0.00		0.00		0.00		
TRIAROMATICS	1.74		1.56		1.26		0.03	
PHENANTHRENES/ANTHRACENE	1.42	1.25		1.00		0.00		
PHENANTHROCYCLOPARAFFINS	0.32	0.30		0.26		0.03		
TETRAAROMATICS	0.21		0.05		0.06		0.00	
PYRENES/BENZOFLUORENES	0.19	0.04		0.05		0.00		
CHRYSENES/BENZANTHRACENE	0.02	0.00		0.00		0.00		
PENTAAROMATICS	0.00		0.00		0.00		0.00	
CHRYSOCYCLOPARAFFINS	0.00	0.00		0.00		0.00		
BENZPYRENES/PERYLENES	0.00	0.00		0.00		0.00		
DIBENZANTHRACENES	0.00	0.00		0.00		0.00		
UNIDENTIFIED	0.00		0.00		0.00		0.00	
Cn.H2n-32	0.00	0.00		0.00		0.00		
Cn.H2n-36/Cn.H2n-26.8	0.00	0.00		0.00		0.00		
Cn.H2n-38/Cn.H2n-28.8	0.00	0.00		0.00		0.00		
Cn.H2n-42/Cn.H2n-32.8	0.00	0.00		0.00		0.00		
Cn.H2n-44/Cn.H2n-34.8	0.00	0.00		0.00		0.00		
AROMATIC SULFUR	0.77		0.58		0.34		0.02	
BENZOTHIOPHENES	0.07	0.58		0.34		0.02		
DIBENZOTHIOPHENES	0.69	0.00		0.00		0.00		
BENZONAPHTHOTHIOPHENES	0.02	0.00		0.00		0.00		

GULF CANADA RESEARCH AND DEVELOPMENT DEPARTMENT

TABLE 4.2.17 MASS SPECTROSCOPY DATA COMPARISON : GULF CATALYST A

	FEED	6012		6014		6013	
RUN #							
HYDROCARBON TYPE	CALCULATED MASS PERCENT						
SATURATES	55.24	82.76		90.67		99.22	
PARAFFINS	11.20	11.75		11.25		13.10	
CYCLOPARAFFINS	44.04	71.02		79.42		86.12	
NONCOND CYCLOPARAFFINS	19.98	28.92		34.32		50.75	
COND DICYCLOPARAFFINS	16.63	29.32		32.38		30.05	
COND POLYCYCLOPARAFFINS	7.43	12.78		12.71		5.32	
AROMATICS	44.76	17.24		9.33		0.78	
MONOAROMATICS	35.05	16.08		8.56		0.77	
ALKYLBENZENES	17.11	10.69		6.65		0.77	
BENZOCYCLOPARAFFINS	11.90	3.74		1.33		0.00	
BENZODICYCLOPARAFFINS	6.04	1.65		0.58		0.00	
DIAROMATICS	6.99	0.21		0.09		0.00	
NAPHTHALENES	3.09	0.21		0.09		0.00	
NAPHTHOCYCLOPARAFFINS	1.66	0.00		0.00		0.00	
FLUORENES	2.24	0.00		0.00		0.00	
TRIAROMATICS	1.74	0.92		0.68		0.00	
PHENANTHRENES/ANTHRACENE	1.42	0.75		0.53		0.00	
PHENANTHROCYCLOPARAFFINS	0.32	0.17		0.15		0.00	
TETRAAROMATICS	0.21	0.01		0.00		0.00	
PYRENES/BENZOFLUORENES	0.19	0.01		0.00		0.00	
CHRYSENES/BENZANTHRACENE	0.02	0.00		0.00		0.00	
PENTAAROMATICS	0.00	0.00		0.00		0.00	
CHRYSOCYCLOPARAFFINS	0.00	0.00		0.00		0.00	
BENZPYRENES/PERYLENES	0.00	0.00		0.00		0.00	
DIBENZANTHRACENES	0.00	0.00		0.00		0.00	
UNIDENTIFIED	0.00	0.00		0.00		0.00	
Cn.H2n-32	0.00	0.00		0.00		0.00	
Cn.H2n-36/Cn.H2n-26.S	0.00	0.00		0.00		0.00	
Cn.H2n-38/Cn.H2n-28.S	0.00	0.00		0.00		0.00	
Cn.H2n-42/Cn.H2n-32.S	0.00	0.00		0.00		0.00	
Cn.H2n-44/Cn.H2n-34.S	0.00	0.00		0.00		0.00	
AROMATIC SULFUR	0.77	0.00		0.00		0.00	
BENZOTHIOPHENES	0.07	0.00		0.00		0.00	
DIBENZOTHIOPHENES	0.69	0.00		0.00		0.00	
BENZONAPHTHOTHIOPHENES	0.02	0.00		0.00		0.00	

TABLE 4.2.18

HYDROTREATING CONDITION OPTIMIZATION

Feedstock: Syncrude Middle Distillate 149-385°C
Catalyst : KATALCO KF506

RUN #	2025	2024	2023
Run Conditions			
Pressure, MPag (psig)	10.3 (1500)	10.3 (1500)	10.3 (1500)
W.A.B.Temperature, °C (°F)	335 (635)	360 (680)	385 (725)
L.H.S.V.	1.0	1.0	1.0
Gas Rate, L/L (SCFB)	890 (5000)	890 (5000)	890 (5000)
H_2 Consumption, L/L (SCFB)	74 (417)	98 (551)	109 (612)
Yields			
Product sg @ 15°C	0.8659	0.8584	0.8504
Feed sg @ 15°C	0.8838	0.8838	0.8838
Vol/vol % yield of TLP on Feed	102.1	103.0	103.9
Product Inspections			
GCD Light Ends, liq. vol. %			
Ethane	nil	0.10	0.06
Propane	0.04	0.19	0.18
IsoButane	0.02	0.05	0.08
N-Butane	0.04	0.15	0.35
IsoPentane	nil	0.05	0.17
N-Pentane	0.08	0.12	0.28
C_6^+	99.8	99.3	98.9
Distillation, D86 °C (°F)			
IBP	157 (315)	139 (282)	80 (176)
10%	203 (397)	199 (390)	187 (369)
20%	224 (435)	219 (426)	206 (403)
30%	243 (469)	238 (460)	226 (439)
40%	254 (489)	255 (491)	242 (468)
50%	274 (525)	269 (516)	258 (496)
60%	287 (549)	283 (541)	272 (522)
70%	302 (576)	297 (567)	287 (549)
80%	316 (601)	312 (594)	302 (576)
90%	332 (630)	330 (626)	322 (612)
FBP	348 (658)	348 (658)	341 (646)
Estimated Molecular Weight	194	194	189
Sulphur, wt ppm	19	<2	<2
Nitrogen, wt ppm	<1	1.6	<1
Naphtha Make, vol% on Feed	0.0	1.1	3.1
Product Composition, mass spec. wt%			
Paraffins	11.6	11.7	12.5
Cycloparaffins	58.1	64.6	65.4
Monoaromatics	28.8	22.6	21.1
Diaromatics	0.6	0.5	0.5
Triaromatics+	0.7	0.6	0.5
Thiophenes	0.3	0.0	0.0

TABLE 4.2.19

HYDROTREATING CONDITION OPTIMIZATION

Feedstock: Syncrude Middle Distillate 149-385°C
Catalyst : KATALCO KF506

RUN #	2020	2021	2022	2032
Run Conditions				
Pressure, MPag(psig)	13.8 (2000)	13.8 (2000)	13.8 (2000)	13.8 (2000)
W.A.B.Temperature, °C (°F)	335 (635)	360 (680)	385 (725)	410 (770)
L.H.S.V.	1.0	1.0	1.0	1.0
Gas Rate, L/L (SCFB)	890 (5000)	890 (5000)	890 (5000)	890 (5000)
H_2 Consumption, L/L (SCFB)	103 (578)	126 (708)	138 (772)	142 (796)
Yields				
Product sg @ 15°C	0.8613	0.8528	0.8430	0.8271
Feed sg @ 15°C	0.8838	0.8838	0.8838	0.8838
Vol/vol % yield of TLP on Feed	102.6	103.6	104.8	106.9
Product Inspections				
GCD Light Ends, liq. vol. %				
Ethane	nil	0.01	0.02	0.08
Propane	0.04	0.07	0.14	0.33
IsoButane	nil	0.03	0.08	0.17
N-Butane	0.06	0.14	0.27	0.62
IsoPentane	0.02	0.05	0.18	0.40
N-Pentane	nil	0.13	0.27	0.62
C_6^+	99.9	99.6	99.0	97.8
Distillation, D86 °C (°F)				
IBP	160 (320)	144 (291)	84 (183)	71 (160)
10%	206 (403)	195 (383)	179 (354)	153 (307)
20%	224 (435)	214 (417)	197 (387)	183 (361)
30%	242 (468)	234 (453)	216 (421)	199 (390)
40%	259 (498)	250 (482)	231 (448)	216 (421)
50%	274 (525)	266 (511)	245 (473)	232 (450)
60%	288 (550)	281 (538)	260 (500)	246 (475)
70%	302 (576)	295 (563)	275 (527)	260 (500)
80%	316 (601)	310 (590)	291 (556)	278 (532)
90%	334 (633)	330 (626)	314 (597)	300 (572)
FBP	347 (657)	344 (651)	335 (635)	328 (622)
Estimated Molecular Weight	195	194	182	172
Sulphur, wt ppm	6.5	<2	<2	<2
Nitrogen, wt ppm	<1	<1	<1	<1
Naphtha Make, vol% on Feed	0.0	1.0	3.0	9.8
Product Composition, mass spec. wt%				
Paraffins	10.9	10.8	11.5	13.9
Cycloparaffins	70.5	76.8	77.7	70.4
Monoaromatics	17.1	11.4	10.0	15.3
Diaromatics	0.0	0.2	0.1	0.3
Triaromatics+	1.0	0.7	0.6	0.0
Thiophenes	0.5	0.1	0.0	0.0

TABLE 4.2.20

HYDROTREATING CONDITION OPTIMIZATION

Feedstock: Syncrude Middle Distillate 149-385°C
Catalyst : KATALCO NM506

RUN #	2029	2028	2027	2031
Run Conditions				
Pressure, MPag (psig)	17.2 (2500)	17.2 (2500)	17.2 (2500)	17.2 (2500)
W.A.B.Temperature, °C (°F)	335 (635)	360 (680)	385 (725)	410 (770)
L.H.S.V.	1.0	1.0	1.0	1.0
Gas Rate, L/L (SCFB)	890 (5000)	890 (5000)	890 (5000)	890 (5000
H_2 Consumption, L/L (SCFB)	81 (452)	127 (715)	150 (840)	165 (927)
Yields				
Product sg @ 15°C	0.8603	0.8511	0.8402	0.8224
Feed sg @ 15°C	0.8838	0.8838	0.8838	0.8838
Vol/vol % yield of TLP on Feed	102.7	103.8	105.2	107.5
Product Inspections				
GCD Light Ends, liq. vol. %				
Ethane	0.01	0.02	0.03	0.08
Propane	0.02	0.07	0.16	0.28
IsoButane	0.03	0.03	0.07	0.17
N-Butane	0.06	0.14	0.28	0.58
IsoPentane	0.05	0.09	0.15	0.42
N-Pentane	0.05	0.12	0.26	0.59
C_6^+	99.8	99.5	99.1	97.9
Distillation, D86 °C (°F)				
IBP	150 (302)	137 (279)	90 (194)	60 (140)
10%	195 (383)	193 (379)	185 (365)	151 (304)
20%	216 (421)	213 (415)	206 (403)	179 (354)
30%	236 (457)	232 (450)	225 (437)	201 (394)
40%	252 (486)	249 (480)	243 (469)	217 (423)
50%	268 (514)	265 (509)	258 (496)	233 (451)
60%	281 (538)	280 (536)	273 (523)	249 (480)
70%	295 (563)	294 (561)	287 (549)	264 (507)
80%	311 (592)	308 (586)	303 (577)	283 (541)
90%	332 (630)	326 (619)	322 (612)	305 (581)
FBP	348 (658)	341 (646)	348 (648)	330 (626)
Estimated Molecular Weight	194	194	190	172
Sulphur, wt ppm	16	10	<2	<2
Nitrogen, wt ppm	<1	<1	<1	<1
Naphtha Make, vol% on Feed	0.5	1.5	2.5	9.9
Product Composition, mass spec. wt%				
Paraffins	10.8	10.7	11.2	14.0
Cycloparaffins	67.3	78.2	82.7	78.6
Monoaromatics	19.5	10.0	5.5	7.2
Diaromatics	0.0	0.1	0.0	0.1
Triaromatics+	1.9	1.0	0.5	0.6
Thiophenes	0.5	0.1	0.1	0.0

TABLE 4.2.21 MIDDLE DISTILLATE HYDROTREATING; KATALCO NM506 M.S. DATA SUMMARY

	2029	2028	2027	2031
	17.2 (2500) 335 (635)	17.2 (2500) 360 (680)	17.2 (2500) 385 (725)	17.2 (2500) 410 (770)
.HYDROCARBON TYPE	CALCULATED MASS PERCENT			
SATURATES	78.12	88.86	93.86	92.59
PARAFFINS	10.81	10.69	11.19	13.96
CYCLOPARAFFINS	67.31	78.17	82.67	78.62
NONCOND CYCLOPARAFFINS	28.16	33.31	38.94	44.94
COND DICYCLOPARAFFINS	26.94	31.97	33.07	29.28
COND POLYCYCLOPARAFFINS	12.21	12.89	10.66	4.41
AROMATICS	21.88	11.14	6.14	7.41
MONOAROMATICS	19.49	9.95	5.50	7.23
ALKYLBENZENES	10.25	6.23	3.02	3.36
BENZOCYCLOPARAFFINS	5.88	2.33	1.47	2.85
BENZODICYCLOPARAFFINS	3.36	1.39	1.01	1.02
DIAROMATICS	0.00	0.09	0.04	0.09
NAPHTHALENES	0.00	0.09	0.04	0.09
NAPHTHOCYCLOPARAFFINS	0.00	0.00	0.00	0.00
FLUORENES	0.00	0.00	0.00	0.00
TRIAROMATICS	0.87	0.89	0.50	0.05
PHENANTHRENES/ANTHRACENE	0.56	0.63	0.37	0.00
PHENANTHROCYCLOPARAFFINS	0.31	0.26	0.13	0.05
TETRAAROMATICS	0.20	0.08	0.04	0.03
PYRENES/BENZOFLUORENES	0.20	0.08	0.00	0.03
CHRYSENES/BENZANTHRACENE	0.00	0.00	0.03	0.00
PENTAAROMATICS	0.81	0.00	0.00	0.00
CHRYSOCYCLOPARAFFINS	0.00	0.00	0.00	0.00
BENZPYRENES/PERYLENES	0.00	0.00	0.00	0.00
DIBENZANTHRACENES	0.81	0.00	0.00	0.00
UNIDENTIFIED	0.00	0.00	0.00	0.00
Cn.H2n-32	0.00	0.00	0.00	0.00
Cn.H2n-36/Cn.H2n-26.S	0.00	0.00	0.00	0.00
Cn.H2n-38/Cn.H2n-28.S	0.00	0.00	0.00	0.00
Cn.H2n-42/Cn.H2n-32.S	0.00	0.00	0.00	0.00
Cn.H2n-44/Cn.H2n-34.S	0.00	0.00	0.00	0.00
AROMATIC SULFUR	0.51	0.13	0.07	0.00
BENZOTHIOPHENES	0.16	0.13	0.07	0.00
DIBENZOTHIOPHENES	0.35	0.00	0.00	0.00
BENZONAPHTHOTHIOPHENES	0.00	0.00	0.00	0.00

TABLE 4.2.22

HYDROTREATING PRODUCTION RUN

Feedstock: Syncrude Middle Distillate 149-385°C
Catalyst : KATALCO NMS06

RUN #	FEED	2027	2035
Run Conditions			
Pressure, MPag(psig)		17.2 (2500)	17.2 (2500)
W.A.B.Temperature, °C (°F)		385 (725)	385 (725)
L.H.S.V.		1.0	1.0
Gas Rate, L/L (SCFB)		890 (5000)	890 (5000)
H_2 Consumption, L/L (SCFB)		150 (840)	158 (885)
Yields			
Product sg @ 15°C		0.8402	0.8388
Feed sg @ 15°C	0.8838	0.8838	0.8838
Vol/vol % yield of TLP on Feed		105.2	106.1
Product Inspections			
GCD Light Ends, liq. vol. %			
Ethane	---	0.03	N/A
Propane	---	0.16	
IsoButane	---	0.07	
N-Butane	---	0.28	
IsoPentane	---	0.15	
N-Pentane	---	0.26	
C_6^+	---	99.1	
Distillation, D86 °C (°F)			
IBP	173 (343)	90 (194)	43 (109)
10%	206 (403)	185 (365)	165 (329)
20%	--- (---)	206 (403)	198 (388)
30%	247 (477)	225 (437)	220 (428)
40%	--- (---)	243 (469)	240 (464)
50%	281 (538)	258 (496)	257 (495)
60%	--- (---)	273 (523)	273 (523)
70%	310 (590)	287 (549)	286 (547)
80%	--- (---)	303 (577)	304 (579)
90%	341 (646)	322 (612)	321 (610)
FBP	357 (675)	348 (648)	347 (657)
Sulphur, wt ppm	1000	<2	<2
Nitrogen, wt ppm	459	<1	<1
TLP Composition, mass spec. wt%			
Paraffins	11.2	11.2	14.3
Cycloparaffins	44.0	82.7	80.0
Monoaromatics	35.1	5.5	5.2
Diaromatics	7.0	0.0	0.0
Triaromatics	1.7	0.5	0.5
Total Aromatics	44.8	6.1	5.7

TABLE 4.2.23 MIDDLE DISTILLATE HYDROTREATING; PRODUCED NAPHTHA COMPOSITION

GULF CANADA RESEARCH AND DEVELOPMENT DEPARTMENT

SAMPLE P12RUN035 P12 RUN 035 NAPHTHA PROJECT 19-84 1984 NOV 9 15:09

INCA & PONA CHARACTERIZATION

INCA (vol%)	17.15	17.00	61.24	4.61
PONA (vol%)	34.15	0.00	61.24	4.61

INCA CHARACTERIZATION BY BOILING RANGE (vol%)

	vol% of sample	Iso-P	n-P	CycloP	Arom
IBP- 40C	10.14	30.47	69.53	0.00	0.00
IBP- 70C	15.87	37.05	60.55	2.40	0.00
IBP-100C	29.18	30.53	39.34	28.07	2.06
IBP-130C	67.24	18.47	19.75	58.06	3.72
IBP-FBP	100.00	17.15	17.00	61.24	4.61

DETAILED COMPOSITIONAL ANALYSIS (vol%)

	PARAFFINS (P)			CYCLOPARAFFINS (C,N)			AROMATICS (A)		
	TOTAL	IsoP	n-P	TOTAL	CP	CH	ALKYLBENZENES	INDANS/ TETRALINS	NAPHTHALENES
C2	0.02		0.02						
C3	0.58		0.58						
C4	3.63	0.65	2.98						
C5	5.91	2.44	3.47	0.38	0.38				
C6	5.35	2.79	2.56	5.95	2.62	3.33	B 0.60		
C7	4.90	3.03	1.87	13.53	3.89	9.64	T 1.90		
C8	5.31	3.51	1.80	19.18	7.59	11.59	X+EB 2.11		
C9	8.45			22.20			0.00	0.00	

* - included in previous value

CALCULATED PROPERTIES

DENSITY, kg/L @ 15 C	0.752	VAPOR PRESSURE, kPa @ 40 C	49.13
AVERAGE MOLECULAR MASS	102.8		

TABLE 4.3.1

CATALYST SCREENING

Feedstock: Syncrude Gas Oil 385+
Catalyst : Katalco KT-550

RUN #	6005B	6005A	6004
Run Conditions			
Pressure, MPag(psig)	20.7 (3000)	20.7 (3000)	20.7 (3000)
W.A.B.Temperature, °C (°F)	385 (725)	405 (761)	420 (788)
L.H.S.V.	0.5	0.5	0.5
Gas Rate, L/L (SCFB)	890 (5000)	890 (5000)	890 (5000)
Gas Make, L/L (SCFB)	--- (----)	--- (----)	--- (----)
Yields (corrected to 100% wt. bal.)			
Product sg @ 15°C	0.9085	0.8790	0.8479
Feed sg @ 15°C	0.9549	0.9549	0.9549
vol./vol. % yield of TLP on Feed	103.7	107.0	108.5
Wt% TLP on Feed	99.8	98.5	96.4
Product Inspections			
GCD Light Ends, liq. vol. %			
Ethane	nil	nil	0.01
Propane	nil	0.05	0.24
IsoButane	nil	0.03	0.15
N-Butane	0.02	0.24	0.72
IsoPentane	0.03	0.17	0.57
N-Pentane	0.07	0.32	0.83
C_6+	99.89	99.19	97.49
Distillation, D86 °C (°F)			
IBP	98 (208)	80 (176)	70 (158)
10%	304 (579)	215 (419)	140 (284)
20%	334 (633)	295 (563)	202 (396)
30%	362 (684)	338 (640)	257 (495)
40%	383 (721)	357 (675)	285 (545)
50%	400 (752)	371 (700)	315 (599)
60%	414 (777)	382 (720)	337 (639)
70%	431 (808)	404 (759)	360 (680)
80%	449 (840)	--- (---)	379 (714)
90%	481 (898)	--- (---)	396 (745)
FBP	--- (---)	--- (---)	--- (---)
(* Distillation by G.C.)			
Estimated Molecular Weight	289	267	208
Sulphur, wt ppm	190	183	28
Nitrogen, wt ppm	113	2.0	1.2
Product Distribution, wt%			
Total Light Ends	0.1	2.1	5.3
Naphtha, C_6-149°C	0.0	0.0	5.0
Mid. Distillate, 149 - 385°C	28.9	45.1	64.0
Gas Oil, 385°C+	71.1	52.9	25.8
TLP (385°C-) Composition, mass spec. wt%			
Paraffins	---	6.5	10.7
Cycloparaffins	---	79.2	79.6
Monoaromatics	---	7.1	6.3
Diaromatics	---	0.7	0.3
Triaromatics+	---	7.3	2.3
Thiophenes	---	1.4	0.8
TLP (385°C+) Composition, mass spec. wt%			
Paraffins	7.9	12.9	9.1
Cycloparaffins	57.2	76.4	82.6
Monoaromatics	17.0	5.2	4.1
Diaromatics	9.8	2.2	1.6
Triaromatics+	4.3	1.1	0.9
Thiophenes	2.3	0.6	0.5
Polars	1.6	1.6	1.3

TABLE 4.3.2

CATALYST SCREENING

Feedstock: Syncrude Gas Oil 385+
Catalyst : GULF CATALYST A

RUN #	6015	6016	6017
Run Conditions			
Pressure, MPag(psig)	20.7 (3000)	20.7 (3000)	20.7 (3000)
W.A.B.Temperature, °C (°F)	385 (725)	405 (761)	420 (788)
L.H.S.V.	0.5	0.5	0.5
Gas Rate, L/L (SCFB)	890 (5000)	890 (5000)	890 (5000)
Gas Make, L/L (SCFB)	70 (390)	143 (804)	265 (1490)
Yields			
Product sg @ 15°C	0.8704	0.8246	0.7856
Feed sg @ 15°C	0.9549	0.9549	0.9549
vol./vol. % yield of TLP on Feed	108.3	110.7	107.4
Wt%, TLP on Feed	98.7	95.6	88.3
Product Inspections			
GCD Light Ends, liq. vol. %			
Ethane	nil	0.04	0.08
Propane	0.12	0.40	0.68
IsoButane	0.14	0.48	1.01
N-Butane	0.31	0.89	1.62
IsoPentane	0.30	0.96	1.92
N-Pentane	0.29	0.88	1.69
C6+	98.84	96.34	92.99
Distillation, D86 °C (°F)			
IBP	76 (169)	51 (124)	43 (109)
10%	160 (320)	116 (241)	97 (207)
20%	244 (471)	159 (318)	125 (257)
30%	309 (588)	196 (385)	147 (297)
40%	346 (655)	232 (450)	172 (342)
50%	365 (689)	261 (502)	196 (385)
60%	380 (716)	298 (568)	217 (423)
70%	391 (736)	331 (628)	244 (471)
80%	402 (756)	350 (662)	275 (527)
90%	--- (---)	386 (727)	326 (619)
FBP	--- (---)	394 (741)	352 (666)
Estimated Molecular Weight	236	180	150
Sulphur, wt ppm	35	2	2
Nitrogen, wt ppm	1.5	<1	<1
Product Distribution, wt% of Feed			
Total Light Ends	2.1	6.9	16.4
Naphtha, C6-149°C	0.0	8.5	21.0
Mid. Distillate, 149 - 385°C	47.2	61.9	58.5
Gas Oil, 385°C+	51.7	22.7	4.4
Product Composition, Mass Spec. wt%			
(149-385°C)			
Paraffins	4.9	11.3	18.6
Cycloparaffins	83.4	85.7	80.4
Monoaromatics	8.5	1.9	1.0
Diaromatics	0.2	0.0	0.0
Triaromatics+	2.7	1.0	0.0
Thiophenes	0.4	0.1	0.0
(385°C+)			
Paraffins	8.0	23.2	36.1
Cycloparaffins	84.0	69.8	62.1
Monoaromatics	3.4	2.6	0.7
Diaromatics	1.9	0.9	0.9
Triaromatics+	1.0	0.5	0.0
Thiophenes	0.5	0.3	0.0
Polars	1.3	2.7	0.2

TABLE 4.3.3

HYDROCRACKING CONDITION OPTIMIZATION

Feedstock: Syncrude Gas Oil 385°C+
Catalyst : GULF CATALYST A

RUN #	6042	6043	6044
Run Conditions			
Pressure, MPag(psig)	13.8 (2000)	13.8 (2000)	13.8 (2000)
W.A.B.Temperature, °C (°F)	360 (680)	385 (725)	400 (752)
L.H.S.V.	0.5	0.5	0.5
Gas Rate, L/L (SCFB)	890 (5000)	890 (5000)	890 (5000)
H_2 Consumption, L/L (SCFB)	51 (288)	63 (355)	135 (756)
Yields			
Product sg @ 15°C	0.9377	0.9308	0.9011
Feed sg @ 15°C	0.9549	0.9549	0.9549
Vol/vol % yield of TLP on Feed	101.8	102.6	106.0
Wt% TLP on Feed	99.8	99.8	98.2
Product Inspections			
GCD Light Ends, liq. vol. %			
Ethane	---	---	---
Propane	---	---	---
IsoButane	---	---	---
N-Butane	---	---	0.10
IsoPentane	---	---	0.07
N-Pentane	---	---	0.15
C_6+	100	100	99.67
Distillation, TBP °C (°F)			
IBP	360 (680)	282 (540)	130 (266)
10%	387 (729)	368 (694)	259 (498)
20%	400 (752)	387 (729)	339 (642)
30%	412 (774)	398 (748)	378 (712)
40%	422 (792)	413 (775)	393 (739)
50%	429 (804)	423 (793)	407 (765)
60%	439 (822)	434 (813)	417 (783)
70%	456 (853)	447 (837)	434 (813)
80%	470 (878)	468 (874)	454 (849)
90%	494 (921)	498 (928)	484 (903)
FBP	530 (986)	530 (986)	521 (970)
Sulphur, wt ppm	118	96	35
Nitrogen, wt ppm	35	19	5
Product Distribution, wt%			
Total Light Ends	0.0	0.0	0.2
Naphtha,C_6 - 149°C	0.0	0.0	1.5
Mid. Distillate,149 - 385°C	6.4	16.8	33.4
Gas Oil,385°C+	93.6	83.2	64.9
TLP (385°C-)Composition, mass spec. wt% *			
Paraffins	--	--	5.5
Cycloparaffins	--	--	53.8
Monoaromatics	--	--	31.8
Diaromatics	--	--	2.0
Triaromatics+	--	--	3.8
Thiophenes	--	--	1.6
TLP (385°C+) Composition, mass spec. wt%			
Paraffins	3.4	3.7	7.5
Cycloparaffins	39.9	44.7	63.9
Monoaromatics	24.4	24.7	14.0
Diaromatics	15.3	13.0	6.6
Triaromatics+	11.0	8.4	3.5
Thiophenes	4.3	3.5	1.5
Polars	1.8	2.0	3.1

* 149°C- removed

TABLE 4.3.4

HYDROCRACKING CONDITION OPTIMIZATION

Feedstock: Syncrude Gas Oil 385°C+
Catalyst : GULF CATALYST A

RUN #	6038	6039	6040
Run Conditions			
Pressure, MPag(psig)	17.2 (2500)	17.2 (2500)	17.2 (2500)
W.A.B.Temperature, °C (°F)	360 (680)	385 (725)	400 (752)
L.H.S.V.	0.5	0.5	0.5
Gas Rate, L/L (SCFB)	890 (5000)	890 (5000)	890 (5000)
H_2 Consumption, L/L (SCFB)	67 (344)	96 (537)	160 (900)
Yields			
Product sg @ 15°C	0.9352	0.9182	0.8894
Feed sg @ 15°C	0.9549	0.9549	0.9549
Vol/vol % yield of TLP on Feed	103.4	104.0	107.6
Wt% TLP on Feed	100.0	99.8	99.2
Product Inspections			
GCD Light Ends, liq. vol. %			
Ethane	---	---	---
Propane	---	---	---
IsoButane	---	---	0.02
N-Butane	---	0.02	0.07
IsoPentane	---	0.07	0.07
N-Pentane	---	---	0.11
C_6+	100	99.91	99.73
Distillation, TBP °C (°F)			
IBP	329 (624)	195 (383)	102 (216)
10%	385 (725)	353 (667)	239 (462)
20%	398 (748)	385 (725)	324 (615)
30%	406 (763)	398 (748)	362 (684)
40%	417 (783)	411 (772)	381 (718)
50%	429 (804)	421 (790)	398 (748)
60%	439 (822)	430 (806)	417 (783)
70%	452 (846)	444 (831)	430 (806)
80%	472 (882)	463 (865)	451 (844)
90%	498 (928)	491 (916)	484 (903)
FBP	530 (986)	530 (986)	517 (963)
Sulphur, wt ppm	90	77	18
Nitrogen, wt ppm	52	19	2
Product Distribution, wt%			
Total Light Ends	0.0	0.1	0.2
Naphtha,C6 - 149oC	0.0	0.0	3.5
Mid. Distillate,149 -385oC	10.2	19.4	39.7
Gas Oil,385oC	89.8	80.5	56.6
TLP (385°C-)Composition, mass spec. wt% *			
Paraffins	--	4.5	6.5
Cycloparaffins	--	51.2	64.4
Monoaromatics	--	29.4	24.0
Diaromatics	--	3.0	1.1
Triaromatics+	--	8.6	2.9
Thiophenes	--	3.3	1.1
TLP (385°C+) Composition, mass spec. wt%			
Paraffins	3.4	6.3	6.3
Cycloparaffins	39.5	56.0	64.0
Monoaromatics	24.3	17.3	17.1
Diaromatics	14.8	10.2	5.8
Triaromatics+	10.3	6.3	3.6
Thiophenes	4.4	2.5	1.5
Polars	3.3	1.4	1.8

* 149°C- removed

TABLE 4.3.5

HYDROCRACKING CONDITION OPTIMIZATION

Feedstock: Syncrude Gas Oil 385°C+
Catalyst : GULF CATALYST A

RUN #	6037	6036	6035
Run Conditions			
Pressure, MPag(psig)	20.7 (3000)	20.7 (3000)	20.7 (3000)
W.A.B.Temperature, °C (°F)	360 (680)	385 (725)	405 (761)
L.H.S.V.	0.5	0.5	0.5
Gas Rate, L/L (SCFB)	890 (5000)	890 (5000)	890 (5000)
H_2 Consumption, L/L (SCFB)	62 (350)	170 (953)	268 (1510)
Yields			
Product sg @ 15°C	0.9322	0.8989	0.8362
Feed sg @ 15°C	0.9549	0.9549	0.9549
Vol/vol % yield of TLP on Feed	102.3	104.7	112.8
Wt% TLP on Feed	99.8	98.3	99.7
Product Inspections			
GCD Light Ends, liq. vol. %			
Ethane	---	---	0.01
Propane	---	0.02	0.11
IsoButane	---	0.01	0.20
N-Butane	---	0.02	0.44
IsoPentane	---	0.05	0.56
N-Pentane	---	0.08	0.56
C_6*	100	99.81	98.12
Distillation, TBP °C (°F)			
IBP	344 (651)	120 (248)	82 (180)
10%	383 (721)	269 (516)	136 (277)
20%	401 (754)	356 (673)	196 (385)
30%	413 (775)	394 (741)	239 (462)
40%	418 (784)	410 (770)	276 (529)
50%	427 (801)	422 (792)	313 (595)
60%	434 (813)	432 (810)	346 (655)
70%	446 (835)	440 (824)	380 (716)
80%	467 (873)	457 (855)	413 (775)
90%	488 (910)	487 (909)	458 (856)
FBP	519 (966)	522 (972)	522 (972)
Sulphur, wt ppm	90	23	3
Nitrogen, wt ppm	11.7	7.3	<1
Product Distribution, wt%			
Total Light Ends	0.0	0.1	1.1
Naphtha, C_6 - 149°C	0.0	3.5	9.5
Mid. Distillate, 149 - 385°C	11.1	27.4	60.7
Gas Oil, 385°C+	88.9	69.0	28.7
TLP (385°C-) Composition, mass spec. wt%*			
Paraffins	4.5	3.7	12.4
Cycloparaffins	43.1	66.0	82.2
Monoaromatics	31.0	23.3	2.9
Diaromatics	8.1	1.2	0.2
Triaromatics+	13.3	5.9	2.3
Thiophenes	4.1	1.2	0.3
TLP (385°C+) Composition, mass spec. wt%			
Paraffins	5.5	5.9	19.5
Cycloparaffins	48.5	68.3	73.1
Monoaromatics	22.1	14.0	3.1
Diaromatics	11.4	5.2	1.0
Triaromatics+	10.0	4.1	1.1
Thiophenes	2.8	1.2	0.4
Polars	2.5	2.5	2.2

* 149°C- removed

TABLE 4.3.6

HYDROCRACKING CONDITION OPTIMIZATION

Feedstock: Syncrude Gas Oil 385°C+
Catalyst : GULF CATALYST A

RUN #	6046	6047
Run Conditions		
Pressure, MPag(psig)	20.7 (3000)	20.7 (3000)
W.A.B.Temperature, °C (°F)	410 (770)	420 (788)
L.H.S.V.	0.5	0.5
Gas Rate, L/L (SCFB)	890 (5000)	890 (5000)
H_2 Consumption, L/L (SCFB)	310 (1741)	392 (2200)
Yields		
Product sg @ 15°C	0.8292	0.7667
Feed sg @ 15°C	0.9549	0.9549
Vol/vol % yield of TLP on Feed	114.2	116.0
Wt% TLP on Feed	97.9	90.6
Product Inspections		
GCD Light Ends, liq. vol. %		
Ethane	---	0.03
Propane	0.23	0.50
IsoButane	0.26	1.08
N-Butane	0.56	1.60
IsoPentane	0.60	2.51
N-Pentane	0.55	2.20
C_6+	97.81	92.08
Distillation, D86 °C (°F)		
IBP	54 (129)	51 (124)
10%	117 (243)	70 (158)
20%	159 (318)	110 (230)
30%	201 (394)	127 (261)
40%	239 (462)	143 (289)
50%	271 (520)	162 (324)
60%	310 (590)	182 (360)
70%	343 (649)	207 (405)
80%	373 (703)	234 (453)
90%	--- (---)	285 (545)
FBP	--- (---)	322 (612)
Sulphur, wt ppm	15	0
Nitrogen, wt ppm	<1	<1
Product Distribution, wt% TLP		
Total Light Ends	1.2	4.4
Naphtha,C_6 - 149°C	13.6	35.4
Mid. Distillate,149 - 385°C	62.3	54.9
Gas Oil,385°C+	22.9	5.3
TLP Composition, 385°C- mass spec. wt% *		
Paraffins	13.6	32.5
Cycloparaffins	80.9	67.4
Monoaromatics	4.0	0.2
Diaromatics	0.3	0.0
Triaromatics+	1.1	0.0
Thiophenes	0.1	0.0
TLP Composition, 385°C+ mass spec. wt%		
Paraffins	18.9	52.9
Cycloparaffins	73.7	40.4
Monoaromatics	4.4	3.3
Diaromatics	1.3	1.1
Triaromatics+	0.6	0.4
Thiophenes	0.3	0.3
Polars	0.8	1.6

*149°C- removed

TABLE 4.3.7

HYDROCRACKING PRODUCTION RUN

Feedstock: Syncrude Gas Oil 385°C+
Catalyst : GULF CATALYST A

RUN #	FEED	6035	6048
Run Conditions			
Pressure, MPag(psig)	20.7 (3000)	20.7 (3000)	20.7 (3000)
W.A.B.Temperature, °C (°F)		405 (761)	405 (761)
L.H.S.V.		0.5	0.5
Gas Rate, L/L (SCFB)		890 (5000)	890 (5000)
H_2 Consumption, L/L (SCFB)		268 (1510)	225 (1264)
Yields			
Product sg @ 15°C		0.8362	0.8487
Feed sg @ 15°C	0.9549	0.9549	0.9549
Vol/vol % yield of TLP on Feed		112.8	111.4
Wt% TLP on Feed		99.7	98.4
Product Inspections			
GCD Light Ends, liq. vol. %			
Ethane		0.01	N/A
Propane		0.11	N/A
IsoButane		0.20	N/A
N-Butane		0.44	N/A
IsoPentane		0.56	N/A
N-Pentane		0.56	N/A
C_6+		98.12	
Distillation, D86/1160°C (°F)			
IBP	396 (745)	82 (180)	78 (172)
10%	408 (766)	136 (277)	132 (270)
20%	--- (---)	196 (385)	200 (392)
30%	419 (786)	239 (462)	240 (464)
40%	--- (---)	276 (529)	275 (527)
50%	437 (819)	313 (595)	314 (597)
60%	--- (---)	346 (655)	345 (653)
70%	458 (856)	370 (698)	370 (698)
80%	--- (---)	392 (738)	392 (738)
90%	501 (934)	--- (---)	--- (---)
FBP	570 (1058)	--- (---)*	--- (---)*
Sulphur, wt ppm	4000	3	6
Nitrogen, wt ppm	1510	<1	<1

* D86 distillation discontinued at 390°C

TLP Composition, 385°C- mass spec. wt% *			
Paraffins	---	12.4	6.4
Cycloparaffins	---	82.2	81.5
Monoaromatics	---	2.9	9.5
Diaromatics	---	0.2	0.4
Triaromatics+	---	2.3	1.8
Thiophenes	---	0.3	0.4
TLP Composition, 385°C+ mass spec. wt%			
Paraffins	4.8	19.5	13.0
Cycloparaffins	36.0	73.1	79.4
Monoaromatics	}	3.1	3.7
Diaromatics	}52.1	1.0	1.3
Triaromatics+	}	1.1	1.1
Thiophenes	2.9	0.4	0.4
Polars	0.6	2.2	1.1

*1490°C- removed

TABLE 4.3.8 GAS OIL HYDROCRACKING; PRODUCED NAPHTHA COMPOSITION
GULF CANADA RESEARCH AND DEVELOPMENT DEPARTMENT

SAMPLE 84-6072-2 NAPHTHA PROJECT 20-84 1984 SEPT 24 10:32

INCA & PONA CHARACTERIZATION

INCA (vol%)	20.73	14.33	61.67	3.27
PONA (vol%)	35.06	0.00	61.67	3.27

INCA CHARACTERIZATION BY BOILING RANGE (vol%)

	vol% of sample	iso-P	n-P	CycloP	Arom
IBP- 40C	8.01	42.82	57.18	0.00	0.00
IBP- 70C	16.48	49.15	47.15	3.70	0.00
IBP-100C	38.06	35.84	28.06	35.00	1.10
IBP-130C	82.00	23.18	15.82	58.79	2.21
IBP-FBP	100.00	20.73	14.33	61.67	3.27

DETAILED COMPOSITIONAL ANALYSIS (vol%)

	PARAFFINS (P)			CYCLOPARAFFINS (C,N)			AROMATICS (A)			
	TOTAL	isoP	n-P	TOTAL	CP	CH	ALKYLBENZENES		INDANS/ TETRALINS	NAPHTH-ALENES
C3	0.06		0.06							
C4	1.72	0.43	1.29							
C5	6.23	3.00	3.23	0.61	0.61					
C6	7.86	4.67	3.19	8.64	5.25	3.39	B	0.42		
C7	8.45	5.54	2.91	18.83	8.78	10.05	T	1.39		
C8	7.66	5.37	2.29	20.13	9.69	10.44	X+EB	1.46		
C9	3.08			13.46				0.00	0.00	

* - included in previous value

CALCULATED PROPERTIES

DENSITY, kg/L @ 15 C	0.746	VAPOR PRESSURE, kPa @ 40 C	31.90
AVERAGE MOLECULAR MASS	100.3		

TABLE 4.3.9 PRODUCT EVALUATION DATA SUMMARY - JET B

	CGSB	H.T. Jet B	H.C. Jet B	Combined* Processing
Freeze Point, °C	-51(max.)	-65	-65	-65
Reid Vapour Press., kpa	21(max.)	5.9	15.9	10.7
Conductivity, pS/m	50(min.) 450(max.)	2	0	2
Distillation, °C				
10%		103	87	95
20%	143(max.)	110	98	108
50%	188(max.)	130	118	128
90%	243(max.)	239	245	240
FBP		281	260	278
% recovered @ 204°C				
Residue, vol%	1.5(max.)	0.2	1.4	
Distillation Loss,	1.5(max.)	1.4	1.4	
Density, kg/L @15°C	0.750(min.) 0.801(max.)	0.7790	0.7689	0.7771
Aromatics, %vol	25(max.)	5.2	4.8	5.1
Olefins, %vol	5(max.)	0	0	0
Sulphur, %mass	0.4(max.)	0	0	0
Sulphur, Mercaptan, %mass	0.003(max.)	0	0	0
Copper Corrosion	No.1(max.)	1	1	1
Water Sep. Index	75(min.)	78	81	78
Net Heat of Combustion MJ/kg	42.8(min.)	43.3	43.2	43.3
Luminometer No.	45(min.)	70.4	72.2	71.0
Thermal Stability, JFTOT colour	3(max.)	1b	1b	1b
pressure drop, kPa	3.4(max.)	0	0	0

TABLE 4.3.10 PRODUCT EVALUATION DATA SUMMARY - JET A

	CGSB	H.T. Jet A	H.C. Jet A	Combined Processing
Flash Point, °C	38(min.)	62	54	62
Freezing Point, °C				
Summer	-40(max.)	-53	-61	-55
Winter	-47(max.)			
Kinematic Vis., cst				
@ -20°C	8(max.)	7.4	6.5	7.2
Distillation, °C				
@ 10%	204(max.)	195	189	189
@ 50%		232	227	229
@ 90%		272	269	270
FBP	300(max.)	281	281	281
vol% residue	1.5(max.)	1.0	0.4	0.5
vol% loss	1.5(max.)	0.8	0.1	0.8
Density, kg/m3	774(min.)			
@ 15°C	839(max.)	840	842	840
Sulphur, mass%	0.2(max.)	0.0	0.0	0.0
Sulphur, Mercaptan %mass	0.003(max.)	0.0004	0.0003	0.0003
Copper Corrosion	No.1(max.)	1b	1b	1b
Silver Corrosion	No.1(max)	0	0	0
Water Sep. Index	75(min.)	82	75	78
Net Heat of Combustion, MJ/kg	42.8(min.)	43.1	42.9	43.0
Aromatics, %vol	22.0(max.)	4.2	6.6	4.6
Luminometer No.	45(min.)	55	49	50
Total Acidity, mg KOH/g	0.1(max.)	0.07	0.04	0.08
Conductivity, pS/m	50(min.)			
	450(max.)	2	12	1
Thermal Stability,				
JFTOT colour	3(max.)	1	1	1
pressure drop	3.4(max.)	0	0.3	0

TABLE 4.3.11 PRODUCT EVALUATION DATA SUMMARY - DIESEL FUEL

	CGSB Type B	H.T. Diesel	H.C. Diesel	Combined Production
Flash Point, °C	40(min.)	69	65	70
Cloud Point, °C		-30	-27	-52
Pour Point, °C		-53	-52	-51
Kinematic Vis., cst. @ 40°C	1.4(min.) 4.1(max.)	2.8	3.3	3.0
Density, g/ml @ 15°C		0.8487	0.8611	0.8598
Distillation, 90% °C	360	328	337	352
Water & Sediment, %vol	0.05(max.)	nil	nil	nil
Total Acid No.	0.10(max.)	0.01	0.01	0.01
Sulphur, %mass	0.7(max.)	0.0	0.0	0.0
Copper Corrosion	No.1(max.)	1	1	1
Carbon Residue on 10% bottoms, %mass	0.15(max.)	0.09	0.10	0.10
Ash, mass%	0.01(max.)	0.00	0.00	0.00
Ignition Quality, Cetane No.	40(min.)	45	46	46
Thermal Stability D2274, mg/100ml		0.9	1.0	0.8